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Full-Scale Dual Alkali FGD (Flue Gas Desulfurization) Demonstration at Louisville Gas and Electric Company

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FULL-SCALE DUAL ALKALI FGD DEMONSTRATION AT LOUISVILLE GAS AND ELECTRIC COMPANY

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16. ABSTRACT The report summarizes the 1-year demonstration of the full-scale dual-alkali flue gas desulfurization (FGD) system at Louisville Gas and Electric Co.'s (LG/E's) Cane Run Unit 6. Systems performance is described in terms of performance guarantees and other parameters that were monitored throughout the demonstration. The report gives a detailed history of operation, including problems encountered in system operation and how they were solved. Capital and operating costs (estimated and incurred) are also reviewed. All performance guarantees (applying to SO2 and particulate matter emissions; lime, soda ash, and power consumption; waste solids properties; and system availability) were met during the demonstration except soda ash consumption and waste solids properties. These two guarantees were not met primarily because of problems related to operation of the vacuum filters. (However, the soda ash consumption guarantee was met during the 12-day acceptance test.) SO2 removal averaged 91.6% during the demonstration and 93.9% during the acceptance test. Monthly performance parameters were applied throughout the demonstration to characterize system operation: FGD system availability was 93.2%; utilization, 77.4%; reliability, 94.1%; and operability, 94.0%.

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Flue Gases	Sodium Carbonates	Stationary Sources	21B
Desulfurization		Dual Alkali System	07A, 07D
Alkalies		Particulates	
Scrubbers			13I
Dust			11G

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The following equivalents can be used for conversion to the metric system:

British	Metric
Btu	1055 J
5/9 (°F-32)	°C
1 ft	0.3048 m
1 ft^3	0.0283 m^3
1 in.	2.54 cm
1 lb (avoir.)	0.4536 kg
1 ton (long)	1.0160 m tons
1 gal.	3.7854 liters
lb/10^6 Btu	430 ng/J
ft/hr	.508 cm/min

ABSTRACT

This report summarizes the one-year demonstration of the full-scale dual alkali flue gas desulfurization (FGD) system, installed at Louisville Gas and Electric Company's (LG&E's) Cane Run Unit 6. System performance is described in terms of performance guarantees and other parameters that were monitored throughout the demonstration period (initially scheduled for 12 months but extended to 13 months). The report presents a detailed history of operation, including problems encountered in system operation and how they were resolved. Capital and operating costs (estimated and incurred) are also reviewed.

The FGD system designer offered LG&E seven process performance guarantees applying to: SO_2 emissions, particulate matter emissions, lime consumption, soda ash consumption, power consumption, waste solids properties, and system availability. All performance guarantees were met during the demonstration period except waste solids properties and soda ash consumption. These two guarantees were not met primarily because of problems related to operation of the vacuum filters. (However, the soda ash consumption guarantee was met during the 12-day acceptance test.) SO_2 removal averaged 91.6% during the demonstration period and 93.9% during the 12-day acceptance test.

Monthly performance parameters were applied throughout the demonstration period to characterize system operation. FGD system availability and utilization were 93.2% and 77.4%, respectively; system reliability was 94.1%, and operability was 94.0%.

The actual capital investment of $19.5 million is somewhat higher than the originally estimated $17.4 million (excluding spare parts and working capital). The material costs for the FGD system and the field supervision are slightly above the original estimate. Material costs for the Conversion Systems, Inc., (CSI) sludge treatment system were nearly double the original estimates (which were based on a different sludge stabilization system), as were the erection labor/overhead costs for the entire system.

Actual annual operating costs for the dual alkali system were much higher than estimated, both in terms of total annual cost and mills per kWh. The estimated total annual operating cost was $5.1 million (with reheat), compared with the $7.8 million actually incurred. The major reasons for this difference were higher labor, maintenance, and soda ash requirements as well as soda ash and fuel oil unit costs that were about twice those estimated.

The dual alkali FGD process was developed by Combustion Equipment Associates, Inc. (CEA, now known as Thyssen-CEA Environmental Systems, Inc.) and Arthur D. Little, Inc. The U.S. Environmental Protection Agency (EPA) selected LG&E's Cane Run Unit 6, a coal-fired boiler of 300-MW peak capacity, as the demonstration plant for dual alkali FGD and participated in funding the operation, testing, and reporting of the project.

The overall demonstration program consisted of four phases: Phase I — preliminary design and cost estimates; Phase II — engineering design, construction, and mechanical testing; Phase III — start-up and acceptance testing; and Phase IV — one-year operation and testing. This report summarizes Phase IV.

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This report was prepared by Arthur D. Little, Inc.; however, the information and data contained represent the work of many individuals from several organizations who have been involved in this project. The principal participating organizations were Louisville Gas and Electric, Inc.; Arthur D. Little, Inc.; and Bechtel National, Inc.

In addition, we would like to acknowledge other efforts and contributions. Norman Kaplan, the EPA Project Officer for this demonstration, has made important technical contributions and has been instrumental in the management of the entire project. Michael A. Maxwell, Director of Emissions/Effluent Technology at EPA's Industrial Environmental Research Laboratory, was responsible for overall planning and review for this program and has provided invaluable guidance and support.

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SECTION 1
INTRODUCTION

1.1 PURPOSE OF PROGRAM

The purpose of this program was to demonstrate the full-scale utility application of dual alkali (D/A) flue gas desulfurization (FGD) technology. The program was a joint effort by several organizations under the sponsorship of the Environmental Protection Agency (EPA). The system was designed by the air pollution control group of Combustion Equipment Associates (CEA)* and Arthur D. Little, Inc. (ADL) and installed on Unit 6 of Louisville Gas and Electric Company's (LG&E) Cane Run Station in Louisville, Kentucky. Unit 6 is an existing 300-MW (gross peak capacity) coal-fired boiler burning high sulfur coal. Installation was by the construction department of LG&E, with technical guidance on process design from CEA.

The FGD system was installed to comply with the requirements of the Jefferson County Air Pollution Control District, the Kentucky State Division of Air Pollution, and Region IV of the U.S. EPA (85% removal of SO_2). EPA selected the D/A process at LG&E as the demonstration system for this technoiogy and participated in funding the operation, testing, and reporting of the project.

The dual alkali system was designed to control SO_2 emissions to less than 200 ppm (dry basis) without additional air dilution when applied to a boiler burning coal containing up to 5% sulfur. When coal containing greater than 5% sulfur is fired, the system is designed to remove at least 95% of the SO_2 in the inlet flue gas. The dual alkali system is not designed for removal of particulate matter. It is designed, however, not to increase the loading of particulate matter in the flue gas.

As a demonstration system, the purpose of the installation and operation was to establish:

- Overall performance — with respect to SO_2 removal, lime consumption, soda ash consumption, regeneration of spent liquor, water balance, scaling or solids buildup, materials of construction, waste cake properties, and system dependability.
- Economics — capital investment and operating cost.

1.1.1 Performance Guarantees

CEA/ADL offered LG&E seven performance guarantees that were passed through in the contract between EPA and LG&E. These performance guarantees, summarized in Table 1 and discussed more fully in Section 4, apply to SO_2 emissions, particulate matter emissions, lime consumption, soda ash consumption, power consumption, waste solids properties, and system availability.

*Now part of Thyssen-CEA Environmental Systems, Inc.

TABLE 1

PERFORMANCE GUARANTEES

Parameter	Guarantee
SO_2 Emissions Limit	200 ppm (dry basis) without additional air dilution (for up to 5.0% sulfur coal)
Calcium Consumption	1.05 moles available CaO per mole SO_2 removed
Soda Ash Consumption	0.045 moles $Na_2 CO_3$ per mole SO_2 removed
Particulate Matter Addition	No net particulate matter addition by FGD System
Power Consumption	System to consume not more than 1.2% of power generated at peak capacity (excluding reheat)
Filter Cake Properties	Filter cake to contain a minimum of 55% (weight basis) insoluble solids
System Availability*	The system shall have an availability (as defined by the Edison Electric Institute) of at least 90% for one year. (The Edison Electric Institute defines operating availability as "the percent of time the unit was available for service, whether operated or not. It is equal to available hours divided by the total hours in the period under consideration and is expressed as a percentage.")

*This guarantee was amended in May 1980 in a contract between EPA and LG&E to require an operability of 90% or greater, coupled with a daily average SO_2 removal efficiency of 90% or greater (see Section 4).

1.1.2 Capital and Operating Costs

Capital and operating cost estimates were developed at the start of the demonstration program. These are summarized in Section 9 and are compared with the actual costs incurred at completion of construction and throughout the demonstration program.

1.2 SCOPE OF WORK

The scope of work for the overall program included the design, construction, start-up, acceptance testing, and one-year operation of a CEA/ADL concentrated mode dual alkali FGD system applied to Unit 6, a 300-MW coal-fired boiler at LG&E's Cane Run Station. The FGD system was designed to treat all the flue gas emitted at the nominal rated capacity (280 MW), with the capability for treating flue gas equivalent to a minimum boiler load of 60 MW and a maximum load of 300 MW.

1.2.1 Four-Phase Program

The overall program was divided into four phases:

- Phase I — preliminary design and cost estimates.
- Phase II — engineering, design, construction, and mechanical testing.
- Phase III — start-up and acceptance testing.
- Phase IV — one year of demonstration operation and performance testing (later rescheduled for 13 months).

LG&E was the prime contractor with overall responsibility for constructing and operating the system. CEA, under subcontract to LG&E, was responsible for the engineering design, for supplying all process equipment, and for providing engineering assistance during start-up and acceptance testing. CEA was also responsible for compliance with all performance guarantees and equipment warranties. ADL, under subcontract to CEA, supplied process engineering support to CEA during design, start-up, and acceptance testing; and gave process assistance to LG&E for system operation during the program. ADL was also responsible for preparing all reports required under the EPA/LG&E contract.

Bechtel National, Inc., under separate contract with EPA, conducted an independent test program to assess system performance with respect to the guarantees offered by LG&E and by CEA/ADL and to monitor and characterize system performance throughout the demonstration program.

This report covers work performed during Phase IV of the program. The objectives of Phase IV were to:

- Characterize system operation over the range of conditions encountered on Cane Run Unit 6, including the effect on system chemistry of substituting commercial lime for carbide lime.
- Monitor system performance relative to design criteria and performance guarantees specified in the LG&E/EPA contract.

- Evaluate the overall technical and economic feasibility/applicability of the system.
- Demonstrate long-term reliability.
- Demonstrate an environmentally acceptable method of waste disposal.

1.2.2 Original and Actual Schedule

The program schedule for all phases of the dual alkali demonstration is given in Figure 1. The overall program, including the demonstration period, was originally scheduled for 40 months (with an additional one month for completion of the final draft report).

Phases I and II were scheduled to begin simultaneously to expedite the overall program. Phase I (preliminary design) was to be completed in five months including preparation of the draft report. Phase II (engineering design and construction) was scheduled for completion 24 months from the signing of the contract.

Construction was completed in March 1979, and the system was initially charged and started up in April 1979. Various problems with the boiler, the FGD system, and the quality and quantity of lime supplied to the system delayed the acceptance testing until after the demonstration period had begun. The acceptance test was successfully conducted in July 1980, when Unit 6 and all FGD system components were fully and simultaneously available, and the boiler could be operated at (or near) full load. The demonstration period officially started on May 1, 1980, and ended on May 31, 1981.

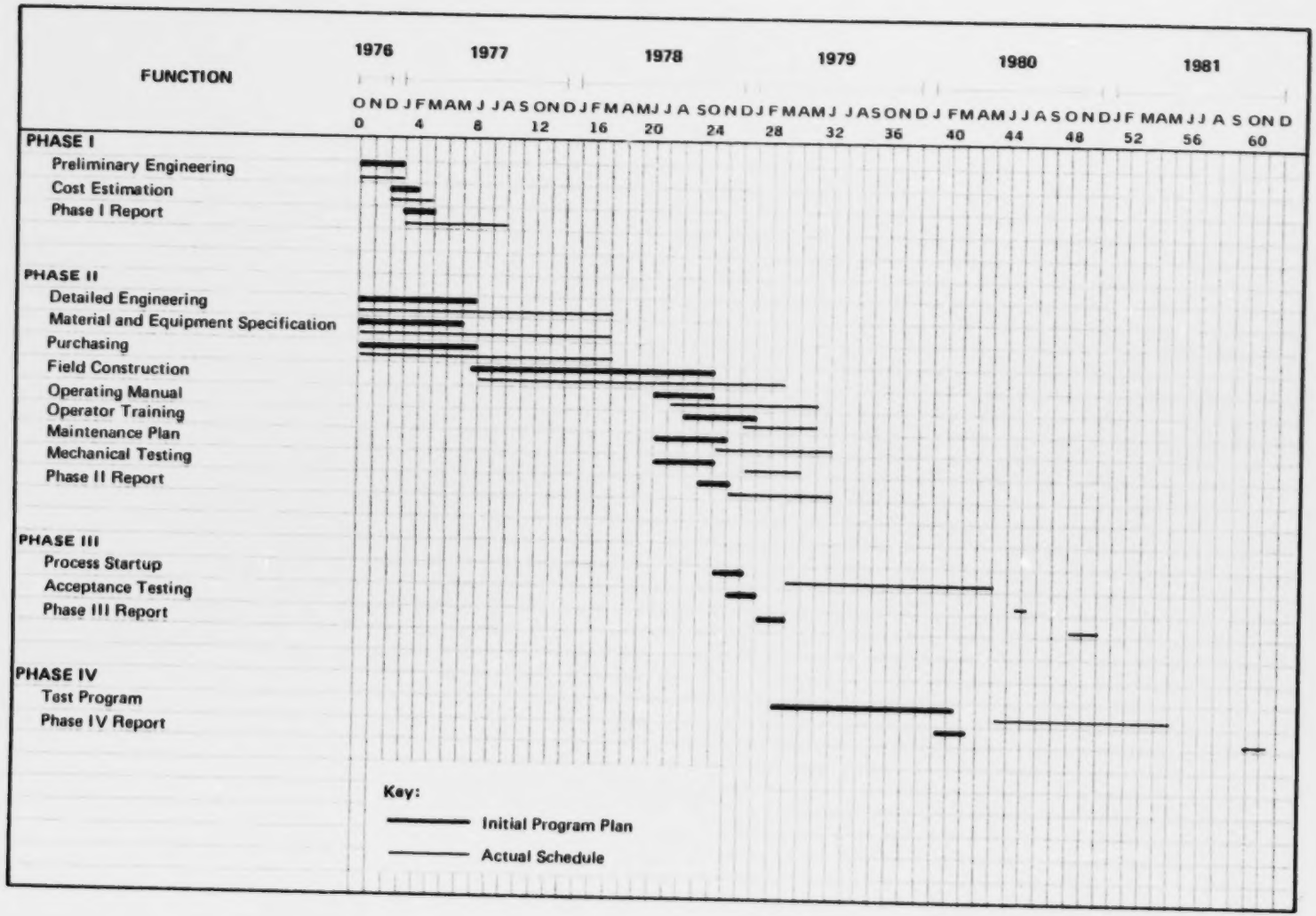
FUNCTION
1976
1977
1978
1979
1980
1981
O N D J F M A M J J A S O N D J F M A M J J A S O N D J F M A M J J A S O N D J F M A M J J A S O N D J F M A M J J A S O N D
0 4 8 12 16 20 24 28 32 36 40 44 48 52 56 60
PHASE I
Preliminary Engineering
Cost Estimation
Phase I Report
PHASE II
Detailed Engineering
Material and Equipment Specification
Purchasing
Field Construction
Operating Manual
Operator Training
Maintenance Plan
Mechanical Testing
Phase II Report
PHASE III
Process Startup
Acceptance Testing
Phase III Report
PHASE IV
Test Program
Phase IV Report
Key:
Initial Program Plan
Actual Schedule

FIGURE 1 SCHEDULE OF DUAL ALKALI DEMONSTRATION PROJECT

SECTION 2
CEA/ADL DUAL ALKALI PROCESS

2.1 GENERAL DESCRIPTION AND PROCESS CHEMISTRY

Dual (or double) alkali is a generic term applied to FGD systems that use soluble alkali to absorb SO_2 and then react the spent scrubber solution with lime and/or limestone to regenerate the alkali and produce a waste calcium-sulfur salt for disposal.

The two principal features of dual alkali technology that distinguish it from conventional direct lime and limestone scrubbing are: (1) the use of a clear solution rather than a slurry for contacting the flue gas in the absorber; and (2) reaction of the solution in a separate absorbent regeneration section to form the waste solids rather than forming the waste solids as a part of the scrubbing operation.

2.1.1 Advantages of Dual Alkali FGD

The dual alkali FGD process was developed to overcome the disadvantages of lime and limestone scrubbing (such as scaling) while retaining the cost advantages of a throwaway system. The use of a scrubbing solution instead of a slurry allows dual alkali FGD to: (1) achieve high SO_2 removal efficiencies over a wide range of inlet SO_2 concentrations (with alkaline solutions); and (2) minimize scale potential in the scrubber circuit. Precipitation of the waste calcium sulfite/sulfate solids outside the scrubber facilitates the formation of waste solids with good dewatering properties. In addition, it is not necessary to wash the mist eliminator to prevent solids deposition and scale formation.

2.1.2 Dilute vs. Concentrated Mode of Operation

Dual alkali FGD systems are usually classified as dilute or concentrated, depending on the concentration of active alkalis in the scrubbing liquor. Active alkalis are ions in the scrubbing liquor that participate in the SO_2 absorption reactions, and they include $CO_3^{=}$, HCO_3^{-}, OH^{-}, $SO_3^{=}$, and HSO_3^{-}. (Although it does not react directly with SO_2, HSO_3^{-} is considered an active alkali because it can be converted to an active form by reaction with lime or limestone.)

The concentration of Na^{+} (sodium ion) in solution that is associated with the active alkali is known as active sodium (although the Na^{+} itself does not participate in the reactions). Dilute systems generally have active sodium concentrations less than 0.15 molar; in concentrated systems the concentrations are greater than 0.15 molar. This distinction is important when determining appropriate methods for sulfate removal. Concentrated systems operate unsaturated with respect to gypsum ($CaSO_4 \cdot 2H_2O$), and gypsum does not precipitate as a separate crystalline phase during regeneration. In concentrated systems, sulfate ions coprecipitate with calcium sulfite during regeneration as a mixed crystal of calcium sulfate and calcium sulfite. In dilute systems, on the other hand, sulfate precipitates during regeneration as gypsum.

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Scrubber oxidation rates are an important factor in concentrated vs. dilute dual alkali systems. Concentrated systems are generally best suited to applications with low oxidation rates — typically those where medium- to high-sulfur coals are burned with little excess air. Dilute systems are preferable where oxidation rates are high. Burning low-sulfur coal promotes high oxidation rates because of the high O_2/SO_x ratio in the flue gas. Boiler firing with excess air has a similar effect. The dual alkali installation at LG&E was designed to operate in the concentrated mode because of the high-sulfur coal used at Cane Run.

2.2 PROCESS SECTIONS

Based on the major steps of dual alkali technology, the FGD system can be broken down into five sections: flue gas scrubbing, absorbent regeneration, solids dewatering, raw materials preparation, and waste disposal.

2.2.1 Flue Gas Scrubbing

In the flue gas scrubbing section, SO_2 is removed as the flue gas is contacted with the alkaline scrubbing solution on the trays of the absorber. The major active constituent of the scrubbing liquor, sodium sulfite, reacts with SO_2 in the flue gas to form bisulfite as follows:

$$Na_2SO_3 + SO_2 + H_2O \longrightarrow 2NaHSO_3 \quad (1)$$

Small amounts of hydroxide and carbonate may also be present in the regenerated scrubbing liquor from the thickener. To reduce the potential for calcium carbonate scale formation on the trays, a portion of the slightly acidic liquor from the bottom of the absorber is blended with the alkaline regenerated liquor from the thickener to maintain a pH of below 10. This slightly acidic recirculated liquor reacts with the carbonate and hydroxide to yield additional sulfite:

$$Na_2CO_3 + 2NaHSO_3 \longrightarrow 2Na_2SO_3 + H_2O + CO_2\uparrow \quad (2)$$

$$NaOH + NaHSO_3 \longrightarrow Na_2SO_3 + H_2O \quad (3)$$

Oxidation of the sodium sulfite to sodium sulfate:

$$2Na_2SO_3 + O_2 \longrightarrow 2Na_2SO_4 \quad (4)$$

must be held to less than 20% by limiting: (1) oxygen concentration in the flue gas or (2) the number of absorber modules operated at low load. Otherwise, the rate at which sulfate can be purged as a mixed crystal will be exceeded, and the solids settling rate will deteriorate as the system approaches dilute mode conditions.

2.2.2 Absorbent Regeneration

The active sodium in the absorber is defined as the sodium salts that can be regenerated for reuse in the scrubbing reaction:

$$[\text{active sodium}] = 2 \times [Na_2SO_3] + [NaOH] + 2 \times [Na_2CO_3] + [NaHSO_3] + [NaHCO_3]$$

The activity of the spent liquor from the absorber is restored by reaction with lime slurry in a two-step process:

$$Ca(OH)_2 + 2NaHSO_3 \longrightarrow CaSO_3 \cdot 1/2H_2O \downarrow + Na_2SO_3 + 3/2H_2O \quad (5)$$

$$Ca(OH)_2 + Na_2SO_3 + 1/2H_2O \rightleftharpoons CaSO_3 \cdot 1/2H_2O \downarrow + 2NaOH \quad (6)$$

Equation 5 represents a neutralization reaction that goes to completion. The equilibrium hydroxide concentration shown in Equation 6 is limited by the relative solubility of calcium sulfite and calcium hydroxide and the concentrations of hydroxide. The reactions shown in Equations 5 and 6 lead to precipitation of calcium sulfite hemihydrate.

A limited amount of calcium sulfate is also precipitated as a mixed crystal of calcium sulfite/sulfate:

$$x\,SO_3^{=} + y\,SO_4^{=} + (x+y)Ca^{++} + zH_2O \longrightarrow Ca_{(x+y)}(SO_3)_x \cdot (SO_4)_y \cdot zH_2O \quad (7)$$

where the ratio of x to y is usually greater than 4:1.

In a concentrated mode, the sulfate coprecipitates with the calcium sulfite, and gypsum (the dihydrate) is not formed. This allows the system to operate unsaturated with respect to calcium sulfate and reduces the potential for gypsum scaling.

The regeneration at Cane Run Unit 6 is accomplished in a two-stage reactor system that prolongs the reactor residence time and produces solids with improved settling and filtering characteristics.

2.2.3 Solids Dewatering

Solids dewatering involves thickening the reactor effluent slurry to 15-30 wt % solids, followed by filtration to produce a moist waste filter cake. While the filter cake is being formed, it is washed with fresh water to recover sodium salts and prevent sodium loss in the process liquor discharged with the cake. The wet filter cake constitutes the only waste discharged from the dual alkali process.

The solids content and chemical composition of the waste cake depend primarily on the amount of oxidation, the quality of lime, the operation of the filter, and the chloride content of the coal. For utility boilers firing medium- to high-sulfur coal, the waste cake for a system operating with commercial lime is expected to contain 55 to 70-wt % insoluble solids and have a chemical composition within the following ranges:

Species	**Weight % (Dry Basis)**
$CaSO_3$	80 to 85
$CaSO_4$	10 to 15
$CaCO_3$ + Inerts	5 to 10
Soluble Sodium Salts	1 to 2

The amount of sodium lost in the cake depends on the sodium concentration in the process liquor, the amount of oxidation, the chloride content of the coal, the operation of the filter, and the permissible level of cake washing.

2.2.4 Raw Materials Preparation

The dual alkali FGD process requires two raw materials: lime for process liquor regeneration and soda ash to compensate for sodium losses. Typically, lime would be hydrated in a slaker before entering the reactor system. At Cane Run, however, carbide lime byproduct is available from a local acetylene plant. Carbide lime is a well hydrated calcium hydroxide slurry containing 25 to 30% solids. The slurry is pumped to a day-tank from which it is fed to the primary reactors. The rate of addition of soda ash to the system is adjusted to compensate for normal losses of sodium in the cake.

2.2.5 Waste Disposal

The FGD waste solids from the filters must be treated to produce a more physically stable and water-impermeable waste for disposal. One such treatment method, the Conversion Systems, Inc. (CSI) process (in which filter cake is blended with about 60% fly ash and 2 to 3% lime based on the dry weight of FGD solids) is used at Cane Run. This waste treatment/disposal system is described in more detail in Section 3.1.

SECTION 3
DESCRIPTION OF THE DUAL ALKALI SYSTEM AT LG&E

3.1 PROCESS DESCRIPTION

Cane Run Unit 6 consists of a pulverized coal-fired steam boiler (built by Combustion Engineering), with a Westinghouse turbine-generator. The unit was designed to operate from a minimum of 60 MW during off-peak hours to a maximum of 300 MW during peak hours. The annual average load is equivalent to approximately 180 MW (about 60% of the gross peak capacity).

Flue gas from the boiler passes in parallel streams through two Ljungstrom combustion air preheaters. Each preheater discharges flue gas through separate ducts to an electrostatic precipitator designed for 99.4% removal efficiency of particulate matter. From the precipitator, the gases enter two parallel induced draft fans, each handling 50% of the total gas.

The SO_2 removal system, installed between the existing induced draft fans and the stack, takes hot flue gas from the outlet of the boiler induced draft fans. Two forced draft booster fans push the flue gas from the ID fans through the absorber. The scrubbed gas from the absorbers is reheated with hot air from an oil-fired burner and then returned to the existing entrance to the stack. In case of scrubber malfunction, dampers have been provided to allow gas bypass around the scrubber system via the existing ductwork.

Coal for Unit 6 is received from several sources. The average sulfur content on a dry basis is approximately 3.5%. (The design for the LG&E dual alkali system is based on 5% sulfur coal.) The average chloride content of the coal is 0.04%, ranging from 0.03% to 0.06%.

3.1.1 Dual Alkali FGD System

The LG&E dual alkali system consists of five process sections:

- Flue gas scrubbing.
- Absorbent regeneration.
- Solids dewatering.
- Raw materials preparation.
- Waste disposal.

Figures 2, 3, and 4 show the process flow diagrams for the overall system.

3.1.1.1 Flue Gas Scrubbing — The flue gas scrubbing section consists of two identical scrubber modules, each containing a booster fan, an absorber, an oil-fired flue gas reheater, and two recirculation pumps (one operating and one back-up).

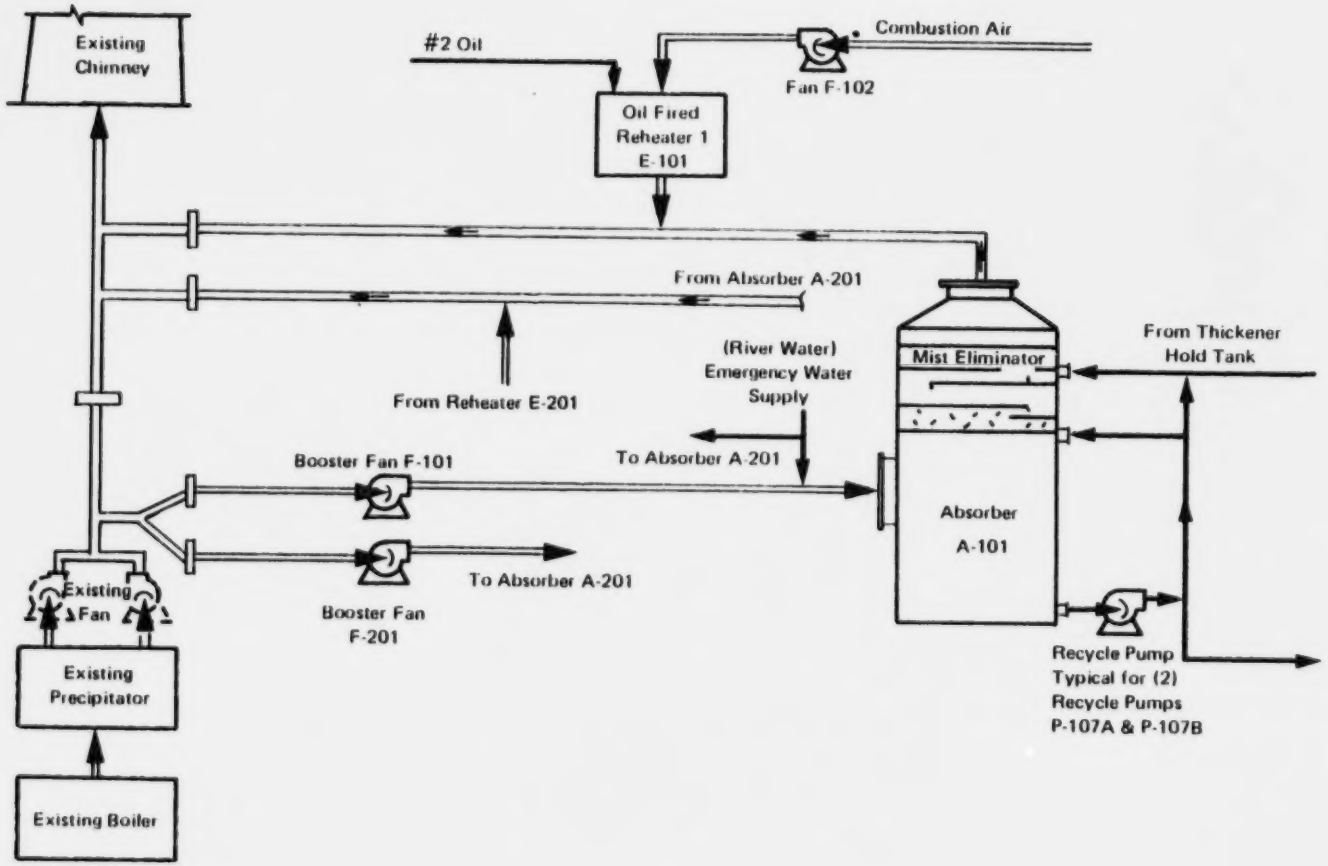

FIGURE 2. LG&E DUAL ALKALI FGD SYSTEM: FLUE GAS SCRUBBING SECTION

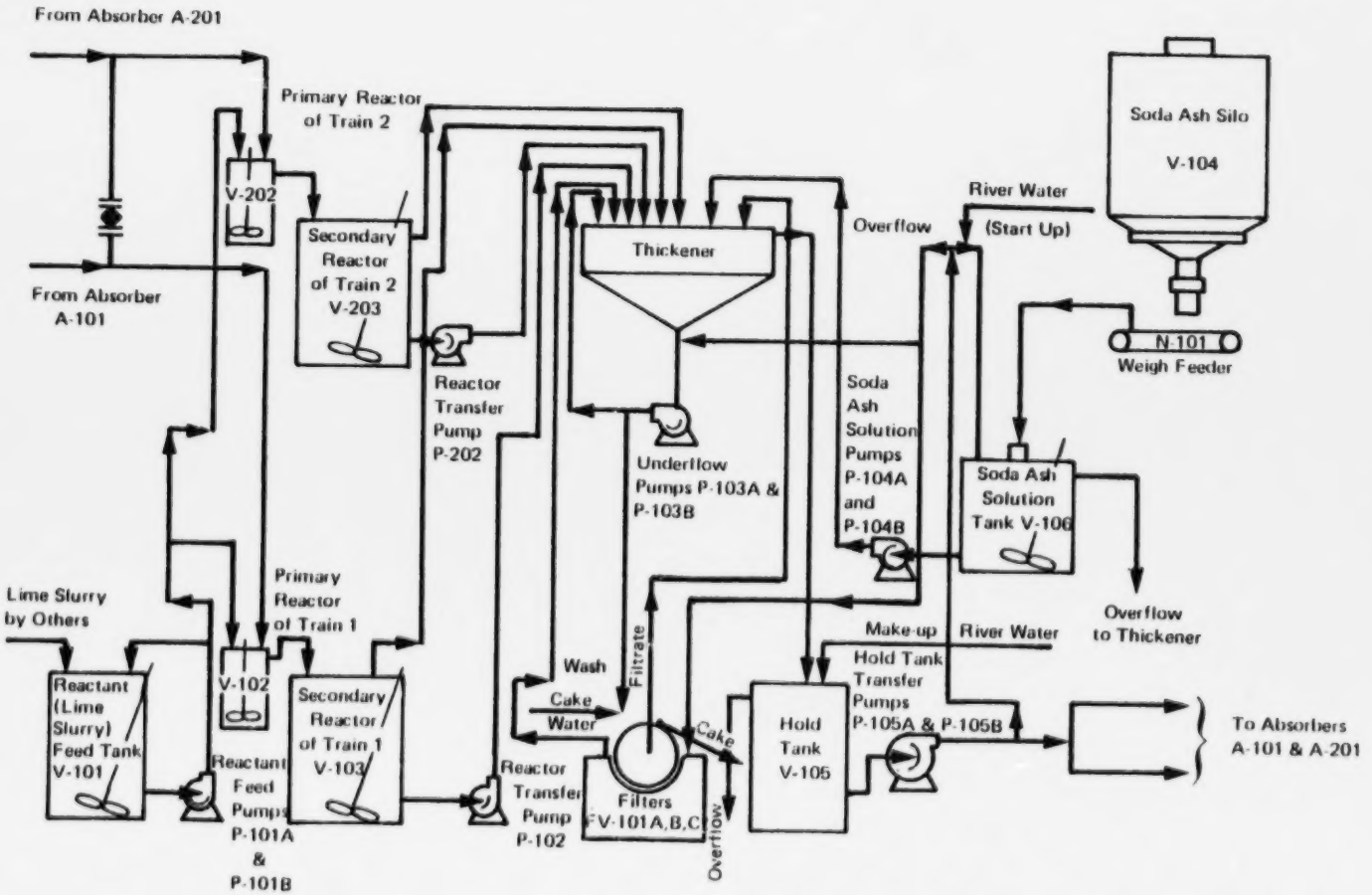

FIGURE 3. LG&E DUAL ALKALI FGD SYSTEM: ABSORBENT REGENERATION SECTION

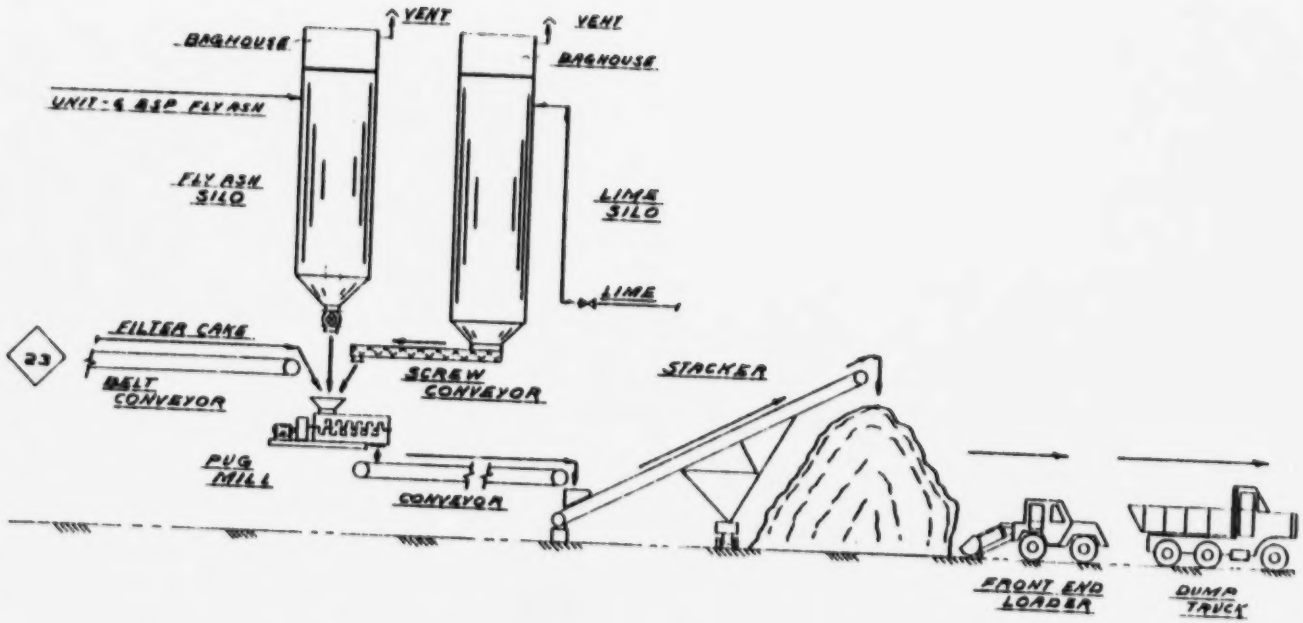

FIGURE 4 LG & E DUAL ALKALI C S I SLUDGE TREATMENT SYSTEM

Flue gas is drawn from the existing induced draft fans and is forced through the 2-stage tray absorbers by the booster fans. Although both absorbers were designed to operate simultaneously (with each handling half the boiler load), one absorber is capable of handling all of the gas during periods of low boiler load. To allow for greater flexibility in operation and provide for the possibility of maintenance on one module while the system is in use, a common duct connecting the two booster fan inlets was incorporated. A bypass was also provided to allow complete shutdown of the scrubbers while the boiler is still on-line. The bypass and FGD system inlet dampers were interlocked to enable the absorbers to be brought on- or off-line without interrupting boiler operation.

Hot flue gas (300°F) entering each absorber is first cooled by a set of sprays that direct scrubbing solution to the underside of the bottom tray. In addition to providing temperature control at the bottom of the absorbers, these sprays minimize accumulation of fly ash solids on the underside of the tray and the bottom of the absorber. The cooled, saturated gas passes through a set of two trays where SO_2 is removed by absorbent liquor that circulates countercurrent to the gas. The flue gas is then routed through a chevron-type mist eliminator to remove any entrained absorber liquor droplets. After leaving the absorber, the scrubbed gas is mixed with hot air from an oil-fired reheater to produce 50°F reheat (to a temperature of about 175°F) to prevent condensation and corrosion.

After the scrubbing solution passes the internal contacting area (two single-pass crossflow perforated plates), it is collected at the bottom of the absorber. Part of the liquor is recirculated to the sprays and part to the trays. Recirculating liquor and regenerated liquor are flow-mixed prior to entering the top tray. The flow rate of the recirculating liquor is adjusted to keep the tray pH below 10. This minimizes the absorption of CO_2 and decreases the potential for $CaCO_3$ scaling in the absorber.

An automatic emergency water supply is provided to quench the hot flue gas and prevent damage to the absorber lining in the event of recycle pump failure. Pressures and temperatures are measured at appropriate points throughout the system. SO_2 removal is monitored by continuous SO_2 and O_2 analyzers at the inlet of the absorbers and in the stack. The pH's of the spent liquor from the absorber and of the top tray feed are also continuously monitored.

3.1.1.2 Absorbent Regeneration — This section of the system consists of two parallel reactor trains. Each contains a primary and secondary reactor. When operating under design conditions (i.e., full load), each reactor train serves to regenerate solution from one absorber. For short periods of time, a single reactor train can handle the solutions from both absorbers operating at full load. When the boiler is firing typical or average coal (3.5% sulfur), only one reactor train is normally required.

The spent scrubbing solution is introduced to the primary reactor along with slurried carbide lime from the lime day-tank. In this reactor (3 to 15 minutes residence time), the regeneration of scrubbing solution begins. The primary reactor overflows into a larger secondary reactor (30 to 60 minutes residence time) where regeneration and crystal formation are completed. The secondary reactor was designed for a pH range of 11 to 12, maintained by controlling the amount of lime slurry fed to the primary reactor.

3.1.1.3 Solids Dewatering — The slurry from the secondary reactor is fed to the thickener feed well to initiate separation of the waste solids from the regenerated solution. The thickener is operated to provide 15 to 30-wt % solids in the underflow, even though the slurry can settle to 40-wt % solids or more. Slurries of about 25-wt % solids allow better control of filter cake washing, control which cannot always be achieved with variations in filter drum speed and vat depth alone. The thickener underflow slurry is recirculated past the filters in a recycle loop that feeds a portion to the filters and returns the remainder to the thickener feed well. Each filter is equipped with an overflow pipe that returns slurry to the thickener to permit operation of the filter in an overflow mode.

There are three filters at LG&E, each rated to handle 50% of the total solids produced at the design conditions. Each filter can be operated independently. For optimum performance (i.e., to obtain cake containing high percentage dry solids and low percentage soluble salts), the filters should run at fixed conditions (constant drum speed, submergence, wash ratio, etc.). Therefore, the cake discharge rate is controlled by changing the number of filters in operation. The number of filters to be placed on line is determined by the quantity of solids accumulated in the thickener. This in turn is judged from the solids concentration in the underflow slurry and the thickener rake lift position indicator. The density of the underflow slurry is measured by a densitometer. (The density of the underflow is monitored by a gamma radiation emission source/detector instrument that measures radiation absorption of the slurry in the pipeline and translates this signal into specific gravity.) Thickener (overflow) hold tank liquor is added at the thickener underflow outlet as required to maintain the solids concentration in the underflow slurry at about 20 to 25%. The number of filters in use is changed if the concentration of solids in the underflow slurry cannot be controlled with the dilution liquor.

The waste cake is washed on the filter by a series of water spray banks. This wash is designed to remove a large fraction (approximately 90%) of the occluded soluble salts from the cake and to return these salts to the system, thereby reducing sodium losses and minimizing sodium carbonate makeup. The total wash rate is usually set at a constant percentage of the cake discharge rate. The mixed filtrate and wash liquor from the filter are collected in the filtrate sump and returned to the thickener. Overall, the total allowed wash rate is controlled by the water balance on the rest of the FGD system.

Clear liquor overflow from the thickener is collected in the thickener hold tank, which provides surge capacity for the absorbent liquor feed to the scrubber system and maintains overall control of the volume of liquor in the system. Water is added to this hold tank to make up for the difference between total system water losses (evaporation and cake moisture) and total water inputs from other sources (pump seals, lime feed, and cake wash).

3.1.1.4 Raw Materials Preparation — Two chemicals are required to operate the dual alkali system at Cane Run Unit 6: carbide lime for liquor regeneration and soda ash to compensate for sodium losses.

Carbide lime, a byproduct of acetylene production, is barged to the Cane Run Station as a slurry containing approximately 25% insoluble solids. The slurry is hydraulically unloaded and pumped to a main storage tank at the plant. Excess liquor is then decanted from the tank and reused to transport the carbide lime. From the main storage tank, the lime slurry is pumped to a classifying and grinding system consisting of a hydroclone and wet ball mill. The fine fraction of lime is then routed to the dual alkali system day-tank, which supplies lime to the primary reactor at the appropriate rate. The coarse fraction is ground in the wet ball mill and sent to lime storage for use in the lime scrubbers on Cane Run Units 4 and 5.

The system was designed to utilize 95% of the available $Ca(OH)_2$ in the carbide lime slurry. Thus the overall feed stoichiometry (moles available $Ca(OH)_2$ per mole of SO_2 absorbed) is expected to be 1.05. The specifications for the carbide lime as delivered to the dual alkali system are given in Table 2.

Dry, dense soda ash is received at the plant and stored in the soda ash silo, which feeds the soda ash solution tank by means of a weigh feeder. The rate of addition is proportional to the amount of lime slurry fed to the system. This relationship exists because the sodium losses in the cake can be expressed as a percentage of the solids produced, which in turn is related to the amount of lime added to the system. The overall rate of soda ash additions is also periodically adjusted based on analysis of the absorbent liquor.

TABLE 2

CARBIDE LIME SPECIFICATIONS*

Constituent	Weight Percent Based on Dry Solids
$Ca(OH)_2$ (equivalent to 70.01% CaO)	92.50
$CaCO_3$	1.85
SiO_2	1.50
$(Fe_2O_3 + Al_2O_3)$	1.60
MgO	0.07
S	0.15
P	0.01
Free carbon	0.25
Not analyzed	2.07

*Available as slurry containing 30% insoluble solids.

Source: Airco catalog (1969)

Soda ash liquor can be made up with either fresh water or clarified liquor drawn from the thickener hold tank. Provisions have been made to feed the soda ash solution either directly to the absorbers or to the thickener. The normal mode of operation is to prepare the soda ash makeup solution using clarified (thickener overflow) liquor and to feed it to the absorbers.

3.1.1.5 Waste Disposal — A long-range disposal plan for the dual alkali filter cake was developed as part of an overall disposal program for all FGD wastes produced at Cane Run. The plan involves stabilization of the wastes generated by each of the three FGD systems via the addition of lime and fly ash, followed by dry landfill of the stabilized material. Waste from the dual alkali system on Unit 6 is handled, processed, and disposed of independently of the wastes produced by the direct lime scrubbing systems on Units 4 and 5. Wastes from these latter two units are combined and handled in a separate common processing plant.

The CSI waste processing plant is installed at the end of the filter building (Figure 4). Filter cake from the dual alkali system discharges into a single conveyor (the three filters are arranged in a single-row, end-to-end), which carries the cake directly to the pug mill mixer in the processing plant. A belt weigh element near the end of the conveyor measures the quantity of cake produced. Fly ash from Unit 6 is pneumatically transported to the processing plant and stored in a live-bottom silo above the pug mill. From there it is discharged through a star valve into the pug mill. The fly ash feed is adjusted according to the filter cake discharge rate and amounts to approximately 60% by weight of the dry filter cake.

Quicklime is also fed to the pug mill at a rate equivalent to 2 to 3% of the dry weight of the cake. The quicklime is stored in a dry silo and fed to the pug mill by screw conveyor.

The pug mill discharges to a radial stacking conveyor. The stacker spreads the material in a windrow (interim stockpiling area) where it remains for 1 to 3 days to initiate the stabilization reactions. The material is then removed from the windrow and loaded into rear dump trucks with a front end loader. The trucks transport the material to a dry landfill on LG&E property south of the plant. Provisions have also been made for direct loading of the trucks from the stacking conveyor.

3.2 PROCESS DESIGN AND CONTROL PHILOSOPHY

The dual alkali system at LG&E was designed to accommodate the characteristics and volume of flue gas to be treated from Cane Run Unit 6 with the types and sources of raw materials available at this location. This section discusses the design basis and control philosophy that were developed for the FGD system.

3.2.1 Design Basis

The design basis for the dual alkali system at LG&E is summarized in Table 3. Additional details of the design basis were presented in the Phase I report for this project (EPA-600/7-78-010).

3.2.2 Water Balance

The LG&E dual alkali system was designed to operate in a closed-loop mode, with no solid or liquid discharge other than that present in the filter cake waste product.

The thickener hold tank was designed to maintain overall control of the volume of liquor in the system. The hold tank collects clear liquor overflow from the thickener and provides surge capacity for the absorbent

TABLE 3

DESIGN BASIS FOR THE DUAL ALKALI SYSTEM AT THE CANE RUN UNIT 6 LOUISVILLE GAS AND ELECTRIC COMPANY

Boiler Conditions	
Boiler Gross Peak Capacity, MW	300
Boiler Gross Net Capacity, MW	277
Annual Load Factor, %	60 (of gross peak capacity)
Btu/kWh (basis-input heat/gross peak electricity generated)	9,960
S in Coal, %	5.0
Cl in Coal, %	0.04
Ash, %	17
Moisture, %	9
Btu/lb of Coal	11,000
Gas Flow Rate @ Scrubber Inlet, acfm	1,065,000
Temperature, °F	300
SO_2, lbs/min.	390
Fly Ash, lbs/min.	4.0
SO_2 @ Scrubber Inlet, lb/10^6 Btu	8.55
SO_2 Removal System	Retrofit
SO_2 @ Scrubber Outlet, lb/10^6 Btu	0.50
SO_2 Removal in Scrubber, %	94.2
Particulate Removal	None
Bypass	Yes
Turndown	to 20% of Capacity
Number of Absorber/Reactor Modules	2
Number of Filter Units	3
Thickeners	1
Carbide Lime Feed System	Existing (new day tank)
Soda Ash Silo Capacity	14 days
Reheat	Direct Reheat with Oil-Fired Reheater
Sludge Disposal	On-site Pond with Treatment[b]
Spare Capacity (Percent)[a]	
Filters	50
Pumps	100
Necessary Instruments to Run Process	100

a. The system can be operated with one reactor train, if necessary, at full load and one absorber system at less than 60% load.
b. Later changed to CSI sludge treatment and landfill.

Source: EPA-600/7-78-010.

liquor feed to the scrubber system. Water is added to the thickener hold tank to compensate for the difference between total system water losses (evaporation and filter cake moisture) and total water inputs from other sources (sodium makeup solution, pump seals, lime feed, and cake wash).

Table 4 shows the overall water balance at design conditions. Process water enters the system in the lime slurry (both as slurry water and chemically combined water), cake wash, instrument purge, and pump seals. (All the water used for pump seals eventually enters the system. The water that drains out of the seals is collected in sumps and returned to the thickener.)

More than 75% of the water lost from the system at design conditions is due to evaporation in the flue gas. The extent of evaporation in the absorbers depends on the boiler load and the temperature and humidity of the inlet flue gas. Water is also removed from the system with the waste cake, both as free water and as chemically bound water.

In addition, water enters and leaves the system via liquid surfaces exposed to the atmosphere (mainly the thickener and thickener hold tank). Based on meteorological conditions at Louisville, thickener and hold tank configurations, and average liquor temperatures, a significant net evaporation of water from the system was anticipated. The annual average evaporation rate from the thickener and hold tank was estimated at 32 gpm, and the average gain from rainfall at about 0.6 gpm.

Since water leaving the system would on the average exceed the uses of water, a net amount of makeup was considered necessary to maintain the system water balance. This was to be added to the thickener hold tank as required to maintain a constant liquid level. The average rate of makeup addition was estimated at 116 gpm for the following design conditions: boiler operating at full capacity (300 MW) and firing 5% sulfur coal.

The process design included the following specific provisions to avoid inadvertent discharge of process water:

- Drains are provided at all pump stations and inside the filter building to direct all process pump seal water leaks, pump and piping flush water, and equipment and building washdown water to the filtrate sump for return to the thickener along with the filtrate and cake wash water.

- A separate sump is provided to collect vacuum pump seal water. This was intended to be non-contact water, but provisions have been made to pump it to the thickener if it becomes contaminated with process liquor.

- The emergency overflow drains from the absorbers are also routed to the filtrate sump for return to the system.

- The thickener has about 3 feet of sidewall above the overflow weir instead of the usual 6 to 8 inches; and the thickener hold tank is as high as the thickener. This increased sidewall height in the thickener and hold tank permits temporary storage of liquor from the other process vessels in the system, if required during maintenance; and also prevents any short-term spills of liquor due to water imbalances resulting from extreme process upsets. The total capacity provided by this additional height equals the total capacity of all the other process vessels in the system.

TABLE 4

OVERALL WATER BALANCE AT DESIGN CONDITIONS

Process Water	**Inlet (gpm)**	**Outlet (gpm)**
Lime Slurry		
— Free Water	128.9	—
— Chemically Combined	12.4	—
Filter Cake Wash	90.7	—
Pump Seals	20.0	—
Instrument Purge	0.5	—
Rain	0.6	—
Make-up (to thickener hold tank)	116.2	—
Evaporation		
— Absorber		278.1
— Open Tanks	—	32.0
Filter Cake		
— Free Water	—	53.1
— Chemically Combined	—	6.1
	369.3	369.3
Non-Contact Water		
Fan Bearing Cooling	175	—
Vacuum Pump Seals	40	—
Fan Bearing Cooling	—	175
Vacuum Pump Seals	—	40
	215	215

- The thickener hold tank is equipped with an emergency overflow tank which starts filling when extreme overflow levels are reached in the thickener and hold tank. Immediate corrective action is necessary when the overflow tank begins to fill.

3.2.3 Control Philosophy

FGD system operating guidelines were developed to facilitate control of the system, especially under varied conditions such as boiler load and coal sulfur content. Instrument redundancy or back-up systems were provided for all major control loops, including fan controls, tank level controllers and indicators, and process pH liquor controllers.

The basic control philosophy for four of the five major process sections — flue gas scrubbing, absorbent regeneration, solids dewatering, and raw materials preparation — is discussed below, according to the various process subsections summarized in Table 5. Waste disposal, the fifth process section, typically involves fixation, such as the CSI process used at Cane Run.

3.2.3.1 Flue Gas Scrubbing — Control and operation of the flue gas scrubbing section, summarized in Table 6, can be conveniently divided as applied to the gas streams and process liquors.

Flue Gas Flow — Control of the flue gas flow to the absorbers is based on maintaining balanced pressure at the boiler I.D. fans. The discharge pressure of the I.D. fans serves as the control parameter. A pressure-indicating controller adjusts the booster fan speed to maintain the pressure in the range of -0.5 to 1.0 inches W.G. (water gauge) (depending upon boiler load). A precalibrated bias is used to send the signal to each booster fan speed controller. The bias adjusts the signal/response to account for mechanical differences in the two booster fan fluid drive units. This effects equal speeds on both fans and provides for parallel fan tracking of boiler load.

Usually both absorbers and fans are operated simultaneously, and each absorber train receives half the boiler load. However, a common duct between the booster fan inlets and appropriate dampers provide the option to operate only one absorber at low boiler load (<60% load). This allows one absorber module to be repaired or cleaned while the system remains in service. It is also desirable to operate with only one absorber module at low boiler load to avoid excessive oxidation of the scrubbing liquor.

Flue Gas Reheat — Heated air/gas is injected into the saturated absorber discharge to provide flue gas reheat. The heated air/gas is generated by diluting combustion gas from a No. 2 oil-fired burner with ambient air. The dilution air mixed with the burner combustion gases is normally maintained at a constant level regardless of load. Control of burner operation is based on the degree of flue gas reheat required, as measured by the differential temperature between the reheated gas and saturated absorber discharge. The differential temperature controls both the burner oil and combustion air fans to provide the preset degree of reheat (usually 50F°).

Emergency Water Sprays — Emergency water sprays on the inlet ducting to each absorber protect the absorber linings from temperature excursions. The emergency sprays are designed to quench the incoming flue gas in the event of a spray system failure. They are activated either by a high temperature reading in the absorber discharge gas or by a low flow rate in the spray recycle line.

Absorber Bleed pH/Feed Forward Flow Rate — The pH of the bleed liquor is an important control parameter for operation of the absorber, since it can be accurately related to the amount of SO_2 removed.

TABLE 5

SUMMARY OF PROCESS SUBSECTIONS OF CANE RUN DUAL ALKALI FGD SYSTEM

Section	Subsections	Variable
Flue Gas Scrubbing	Gas:	Flue Gas Flow
	Flue Gas Reheat:	Oil Flow
	Process Liquor:	Feed Forward (Regenerated Liquor) Flow
		Tray Recycle Flow
		Spray Recycle Flow
		Absorber Bleed Flow
Absorbent Regeneration	Reactor Trains:	Primary Reactor Bleed Flow
		Secondary Reactor Bleed Flow
		Lime Feed Rate
Solids Dewatering	Slurry Thickening:	Thickener Underflow
		Dilution Liquor Flow
		Makeup River Water Flow
	Dewatering:	Drum Speed
		Drum Submergence
		Wash Water Flow
Raw Materials Preparation	Carbide Lime:	Lime Feed
		Lime Recycle
	Soda Ash:	Soda Ash Feed
		Hold Tank Liquor Flow
		Soda Ash Solution Flow
Waste Solids Treatment and Disposal	Fixation:	Fly Ash Addition
		Lime Addition
		Stacking
	Disposal:	Transport
		Landfill

TABLE 6

CONTROL PHILOSOPHY FOR THE FLUE GAS SCRUBBING SECTION

Controlled Variable	Approximate Variable Range (per absorber)	Control Function	Control Parameter	Parameter Range
Gas:				
Flue Gas Flow	106,000 to 533,000 acfm	Gas Flow Balance	Fan Suction Pressure	+ 1.0 to -0.5" W.G.
Reheater Oil Flow	0 to 3.0 gpm	Reheat Scrubbed Gas	Gas Differential Temperature	0 to 50°F
Process Liquor:				
Feed Forward Flow	280 to 1850 gpm	SO_2 Removal	Bleed Liquor pH	6 to 7 gpm
Tray Recycle Flow	260 to 870 gpm	Top Tray pH and Tray Loading at Low Flow	Flow	260 to 870 gpm
Spray Recycle Flow	1600 gpm	Gas Quench	Manual	—
Absorber Bleed Flow	260 to 1700 gpm	Flow Balance	Absorber Tank Level	4' to 5'

Source: EPA-600/7-79-221b

The feed forward flow rate of regenerated scrubbing liquor to the absorber is adjusted to maintain the bleed liquor pH within the range needed to give the required SO_2 concentration in the exit flue gas. The bleed pH was initially controlled at 6, which was estimated to result in an SO_2 concentration in the exit gas of less than 200 ppm. The actual correlation between pH and SO_2 removal was determined during the early periods of operation, and the actual set points were modified accordingly. There is an option for operating the absorber feed forward rate on flow control, with manual adjustment of the flow set point. This is sometimes required during maintenance on the bleed liquor pH monitors.

Tray Recycle — The tray recycle is required to insure a proper liquid loading of the trays because the feed forward rate varies. The tray recycle also controls the pH of the total liquor fed to the top tray and can thus minimize CO_2 absorption and the potential for $CaCO_3$ precipitation. The tray recycle flow rate is normally set at 650 gpm (range of 260-870 gpm). The pH of the tray feed (combined tray recycle and feed forward) normally fluctuates within the range of 7.5 to 10.5. The set point for the tray recycle flow rate may be changed to maintain the pH within this range, although this adjustment is normally required only if the pH falls outside the range for extended periods of time (a few hours or more). Hence, the pH of the combined liquor feed to the top tray is monitored and alarmed but is not used for automatic control of the recycle flow. The flow is maintained at a constant rate and is adjusted manually as required. Increasing the recycle flow rate lowers the pH of the combined stream; decreasing the recycle flow rate has the opposite effect.

Spray Recycle — The spray recycle is important because it serves to quench the incoming flue gas, protect the absorber lining, and wash the bottom tray. The sprays are operated at a constant flow rate of 1,600 gpm (independent of boiler load). The flow rate is manually controlled as required by adjusting the spray recycle valve.

Absorber Bleed Stream — The flow rate of the bleed stream is controlled by the liquid level in the absorber tank, normally maintained at a height of 4.5 ft (range 4 to 5 ft). The flow rate is expected to vary within 260 to 1,700 gpm over the full range of boiler load and coal sulfur content. During maintenance on the absorber level control loop, the absorber bleed rate can be operated on flow control, with manual adjustment of the flow set point.

3.2.3.2 Absorbent Regeneration — Optimal operation of the regeneration system involves controlling the extent of scrubbing solution regeneration to maximize lime utilization and produce waste solids with good dewatering characteristics. The pH of the secondary reactor bleed stream measures the extent of regeneration and thus becomes the controlling factor. Table 7 summarizes the control philosophy for the absorbent regeneration section.

Tank Level/Effluent Flow of Primary Reactors — The absorber bleed streams are mixed with carbide lime in the primary reactors and then overflow into the secondary reactors. The tank level in the primary reactors is thus constant at the overflow level, and the effluent flow rate is the same as the combined feed rate of absorber bleed and carbide lime slurry. The hold-up time in these reactors is dictated by the flow rate of the combined feed streams. The normal operating range for the hold-up time is 3 to 10 minutes.

Tank Level/Effluent Flow of Secondary Reactors — Each secondary reactor is provided with one reactor transfer pump and an overflow line. The effluent (or bleed) from the secondary reactors can be fed to the thickener in one of two ways: the effluent can be pumped (on level control), or the tank can be allowed to operate on overflow. In either case, the secondary reactors are normally operated at hold-up times greater than 30 minutes.

TABLE 7

CONTROL PHILOSOPHY FOR THE ABSORBENT REGENERATION SECTION

Control Variable	Variable Range (per reactor train)	Control Function	Control Parameter	Parameter Range
Primary Reactor Bleed Flow	270-1800 gpm	Reaction Time (4-30 min.)	Tank Level	12 ft
Secondary Reactor Bleed Flow	270-1800 gpm	Reaction Time (40-260 min.)	Tank Level	18-30 ft
Lime Feed Rate	30-120 gpm	Solid Waste Properties/ Lime Utilization	Secondary Reactor Bleed Liquor pH	11-12.0 for Commercial Lime 8.5-9.5 for Carbide Lime

Source: EPA-600/7-79-221b

Under conditions typical of boiler load and coal sulfur content, the secondary reactors are usually operated on overflow (level of 30 ft). The secondary reactors can be operated at lower liquid levels, if necessary, by means of the reactor transfer pumps (with flow controlled on liquid level). In this manner the secondary reactor hold-up times can be adjusted during sustained periods of extremely low boiler loads. The pumps are also used to drain the reactors when they are removed from service for maintenance.

Bleed pH/Lime Slurry Feed — The amount of lime fed to the primary reactor of each train is controlled by the pH of the corresponding secondary reactor bleed stream. The bleed pH is automatically controlled within the range of 8.5-9.5 with carbide lime slurry or 11-12.0 with commercial lime. (When carbide lime is used to regenerate the scrubbing liquor, the operating pH of the secondary reactor must be reduced to prevent aluminum/silica dissolution from the carbide lime and subsequent scaling in the absorbers.)

There are occasions, however, when it is desirable to override pH control to prevent lime overfeeding. This can occur, for example, when there is a rapid and significant change in boiler load or coal sulfur content, or a rapid drop in secondary reactor pH due to some upset condition. Under such circumstances, lime could be overfed to the primary reactors for some period of time, since the lag time for pH sensors to detect the effect of lime addition is greater than one-half hour (primary plus secondary reactor hold-up times). A maximum lime feed rate controller has been provided to avoid this problem. It limits the lime feed rate, which is directly related to the absorber bleed stream flow.

One vs. Two Reactor Train Operation — The number of reactor trains in use is generally dictated by the boiler load and the coal sulfur content. Only one reactor train is usually required with typical coals (~3.5%S), and the secondary reactor is operated on overflow. With high sulfur coals (>4% sulfur) and high sustained boiler loads (>70%), both reactor trains should be operated. As a general rule, the number of reactor trains in service should be the same as the number of absorbers in operation.

3.2.3.3 Solids Dewatering — Separation of the solid wastes from the regenerated scrubbing solution is accomplished in a two-step operation: thickening of the slurry followed by vacuum filtration. Table 8 summarizes the control philosophy for the solids dewatering section.

Slurry Thickening/Thickener — The secondary reactor effluent fed to the thickener contains 2 to 5% solids. The thickener is generally operated to provide a slurry underflow of 20 to 25% solids. This solids concentration permits reasonable control of the filter operation; higher concentrations can create filter plugging problems. Since the solids generated in the reactor system will normally thicken to a much higher solids level (up to 40% or more, if not controlled), clarified liquor from the hold tank is used to control the solids level in the thickener underflow. A density sensing element, located at the discharge of the underflow pumps, measures the slurry concentration and adjusts the rate of dilution liquor addition to maintain a specific gravity consistent with 20 to 25% solids. Part of the slurry is recirculated past the filters and returned to the center well of the thickener. This maintains flow through the lines and prevents solids deposition. The underflow and recycle flow rates are set manually by opening the corresponding valves.

Slurry Thickening/Thickener Hold Tank — The clarified liquor from the thickener overflows into the hold tank. Makeup river water, required to maintain the water balance in the system, is fed to the hold tank on level control. The hold tank liquor, which is regenerated scrubbing solution, is pumped to the absorbers. As indicated previously in the absorber section, the flow rate of the feed forward (hold tank liquor) to each absorber is controlled by the pH of the absorber bleed stream.

TABLE 8

CONTROL PHILOSOPHY FOR THE SOLIDS DEWATERING SECTION

Slurry Thickening	Variable Range	Control Function	Control Parameter	Parameter Range
Thickener Underflow Recycle Flow Rate	250-600 gpm	Solids Deposition	—	—
Dilution Liquor Flow Rate	0-260 gpm	Suspended Solids Concentration	Underflow Specific Gravity	1.20-1.26
Makeup River Water Flow Rate	50-150 gpm	System Water Balance	Hold Tank Level	8-16'
Dewatering				
Drum Speed	0.8-1.2 rpm	Filtration Rate	Cake Thickness	1/4-1/2"
Drum Submergence	8-10"	Filtration Rate	Cake Thickness	1/4-1/2"
Wash Water Flow Rate	40-80 gpm*	Na^+ Losses		

*Per filter

Source: EPA-600/7-79-221b.

Vacuum Filtration/Filter Units — The most important factors in assessing filter performance are the solids content and the level of sodium salts contained in the filter cake (i.e., solids dewatering and washing). These are primarily a function of the quality of the solids produced in the reactor system, the thickness of the filter cake, and the amount of cake washing. In general, optimal performance is achieved with a cake thickness of about 3/8 inch (a range of 1/4 to 1/2 inch) and a wash ratio in the range of 1.5 to 3.0 (gal. of wash water/gal. of water in the filter cake). The wash water rate, therefore, is a function of the rate of discharge of wet cake and the moisture (or solids) content of the cake. Since the solids content varies only slightly and the drum speed is usually fixed, the wash water flow rate can be preset as long as cake thickness is controlled in the proper range. If necessary, it can be readjusted based on the weight of cake produced as measured by the conveyor belt weigh element. The wash flow rate to each filter is initially set at 80 gpm.

The cake thickness is controlled by four parameters, three of which are preset: drum speed (1.0 rpm), drum submergence (8 to 10 inches), and the filter bridge valve position (6 o'clock).

The position of the filter bridge valve determines the time (position) at which vacuum is applied to the filter drum while it is submerged in the filter tub and thereby influences cake thickness. The fourth parameter, the underflow slurry concentration, is used as the primary control parameter to maintain proper cake thickness. As indicated in the thickener section, the slurry concentration is kept at a suspended solids level of about 20 to 25%. However, when the slurry concentration deviates appreciably from this range and cannot be adjusted further by added dilution liquor, an operator must either put additional filters in service or take them off-line.

Vacuum Filtration/Number of Filter Units in Service — The number of filters in operation is dictated by the thickener underflow slurry concentration and the amount of diluting hold tank liquor added, as explained in the previous paragraph. A filter must be taken off line whenever the dilution liquor has been reduced to its minimum rate and the slurry concentration is below about 15%. Similarly, when the dilution liquor rate is at its maximum and the slurry concentration exceeds about 28%, another filter must be placed in service.

Vacuum Filtration/Filtrate Sump — The filtrate is collected in filtrate receivers and directed to the filtrate sump. From there it is pumped to the thickener.

3.2.3.4 Waste Solids Treatment and Disposal — The waste solids from the vacuum filter system are treated prior to landfill disposal to produce a waste both low in permeability and moderate in compressive strength. The low permeability is required to prevent leaching of salts from the waste filter cake into the groundwater and compressive strength is required to support the weight of vehicles or structures on the completed landfill.

Waste Solids Treatment — Wet filter cake is discharged from the vacuum filter onto a belt conveyor which carries it to the CSI system. In the pugmill (paddle wheel mixer) of the CSI system, the wet filter cake is blended with fly ash from the Unit 6 electrostatic precipitator (ESP) and quicklime from a lime storage bin. The ratios of fly ash and quicklime added to the wet filter cake are approximately 0.6 parts fly ash and 0.02-0.03 parts quicklime by weight per part of dry solids in the filter cake.

Treated filter cake from the CSI system is stacked in piles where it remains for 1-3 days for initiation of the cementation reaction.

A filter cake solids content of 50-55% is required for the most efficient handling (mixing) properties in the CSI system, while the moisture content of the treated (stacked) waste filter cake typically ranges from 32-35% (65-68% solids). The use of quicklime rather than hydrated lime helps to minimize moisture content in the stabilized sludge. The ratio of fly ash to filter cake solids is limited by the ratio of sulfur to ash in the coal. At full load, Unit 6 produces 16 tons per hour of fly ash. This amount of fly ash is used in the CSI system for each 48 tons of scrubbing filter sludge (at 55% solids). The quantity of quicklime added to the stabilized waste can be varied, but because it is relatively expensive, its addition is held to the minimum that will yield the desired properties.

Disposal of Treated Waste Solids — A front end loader places the treated filter cake into dump trucks for transport to the landfill site at Cane Run. The waste solids are dumped, spread in 1-1/2 to 2 ft. lifts, and compacted by a vibrating roller. When the height of the filled waste reaches approximately 30 ft. (about 15 lifts), the landfill area will be covered with 18" of soil and seeded.

The stabilized waste has a typical compressive strength of 250-350 psi and a permeability of less than 10^{-7} cm/sec.

3.2.3.5 Raw Materials Preparation — The dual alkali process requires two raw materials: carbide lime and soda ash makeup. Table 9 summarizes the control philosophy for the raw materials handling and feed systems.

Carbide Lime/Lime Feed to Day Tank — Carbide lime is available to LG&E as a slurry containing 25 to 30% insoluble solids. The lime is stored in a day-tank that is filled batchwise as the slurry level drops below the preset point. Initially, the preset level corresponds to about 60% of the tank capacity and requires one or two fillings per shift. The preset level may be changed, however, and the contents of the lime tank may drop as low as 25% of the tank capacity before refill.

Carbide Lime/Lime Tank Recycle — The amount of lime added to the primary reactors is controlled by the pH of the corresponding secondary reactors (with override limitation on maximum feed based upon absorber bleed flow, as discussed previously). The lime slurry is recirculated around the lime feed tank to maintain sufficient flow through the pipes to prevent solids deposition. A bleed from this recycle is fed to the primary reactors. A control valve on the day-tank return line is interconnected with the control valves on the lime slurry feed lines to the individual reactors. The recycle valve maintains enough pressure in the feed lines to ensure sufficient slurry feed and adequate feed control. The valve position on the slurry return line to the day-tank is controlled by the larger of the flows to each of the primary reactors. As the larger of the reactor feed rates increases (decreases), the control valve on the return line closes (opens) to maintain the feed line pressure consistent with a reasonable range of operation for the feed control valves.

Soda Ash/Silo — Dense soda ash is received and stored in the soda ash silo, which has a capacity of approximately 140 tons. Maintenance of the silo inventory normally requires two to four 20-ton trucks per week, depending on the system demand.

Soda Ash/Solution Tank — A weigh feeder is used to feed the soda ash to the soda ash solution tank. The amount fed can be set manually and readjusted daily or weekly as required; or it can be controlled automatically in proportion to the amount of lime fed to the reactor system. (The amount of lime fed dictates the amount of waste cake produced and ultimately the rate of soda ash lost in the washed cake.) On automatic control, the bias in the controller is adjusted periodically (about once a week) to maintain a constant thickener overflow liquor specific gravity. If the specific gravity is increasing, the soda ash should be decreased. Similarly, if the specific gravity is decreasing, the soda ash feed should be increased.

TABLE 9

CONTROL PHILOSOPHY FOR THE RAW MATERIALS PREPARATION SECTION

Control Variable	Variable Range	Control Function	Control Parameter	Parameter Range
Carbide Lime				
Lime Feed	Batch	Lime Inventory	Lime Tank Level	15'-21'
Lime Recycle	150-285 gpm	Solid Deposition	Sufficient Flow for Solids Suspension	150-285 gpm
Soda Ash				
Soda Ash Feed	40-80 tons every week	Soda Ash Inventory	Soda Ash Silo Level	12'-30'
Hold Tank Liquor Flow	30-60 gpm	Sufficient Flow to Dissolve Soda Ash	Flow Rate	30-60 gpm
Soda Ash Solution Flow	30-60 gpm	Soda Ash Solution Feed	Soda Ash Tank Level	4'-6'

Soda ash solution is normally prepared with clarified liquor from the thickener hold tank. Solution liquor from the hold tank is pumped (from the discharge of the hold tank pumps) to the soda ash solution tank at a constant rate, nominally 50 gpm. The resulting soda ash solution is pumped to the absorbers or to the thickener on level control.

3.3 ORIGINAL AND REVISED OPERATING CONDITIONS

The original design and operating conditions for the system, as shown in Table 10, specified pHs of:

- 6.0 for the absorber,
- 11.5 for the secondary reactor,
- 12 for the thickener overflow, and
- 7 to 10.5 for the top tray feed.

Unfortunately, these conditions were ideal for dissolving aluminum in the reactors and thickener and then precipitating the alumina-silica complex on the absorber trays. As a result the operating pH conditions were changed from those originally specified to those shown in Table 11 and summarized below:

- 6.3 for the absorber spent liquor,
- 8.0 for the primary reactor,
- 9.0 for the secondary reactor, and
- $<$10.5 for the thickener overflow.

The alumina dissolution/precipitation problem is discussed in more detail in Section 6.1.2.2.

The absorber liquor pH was raised to compensate for the lower pH of the thickener overflow fed to the top tray (and to maintain SO_2 removal efficiency). This was accomplished by increasing the "feed forward" flow of liquor from the thickener to the absorber.

TABLE 10

INITIAL SET POINTS FOR SCRUBBER OPERATION

	Normal Setpoint	Allowable Range
Absorbers		
Bleed pH	6.0	6.0-6.4
Recycle to Trays (gpm)	650	600-800
Top Tray Feed pH	NA	7-10.5
Reheaters (T)	50°	40°-50°
Number of Modules in Service	(1) < 120 MW (2) > 120 MW	
Reactors		
Lime Slurry Feed Solids	25%	20%-30%
Secondary Reactor pH	11.5	11.0-11.8
Thickener Overflow pH	12.0	11-12
Number of Operating Trains (assumes 1750 ppm to 2500 ppm inlet SO_2)	(1) < 150 MW (2) > 150 MW	
Filters		
Drum Speed (setting)	2-1/2	0.7 rpm to give 3/8" cake
Wash Water (gpm)	60-70 gpm	not less than 60 gpm
Other	No wash nets Not more than 1-3 minor cloth holes Two filters operating min. of 16 hrs each day	
Soda Ash		
Makeup Rate	To maintain Na + Active = 0.40-0.55 M	
Makeup Solution	Use hold tank liquor	
Makeup Flow	~ 50 gpm	
Makeup Feed Location	Thickener	

Source: Arthur D. Little, Inc., memorandum to LG&E dated 11/29/79.

TABLE 11

DESIGN AND REVISED OPERATING CONDITIONS FOR LG&E DUAL ALKALI SYSTEM

Component	Design Operating Conditions	Revised Operating Conditions
Absorber		
Flow Rate to Absorber	280 to 1850 gpm	
Spent Liquor pH	6.0	6.3
Top Tray Feed pH	7.5 to 10.5	
Primary Reactors		
Flow Rate	260 to 1700 gpm	
Retention Time	3 to 15 minutes	
pH	5 to 11	8.0
Secondary Reactors		
Flow Rate	260 to 1700 gpm	
Retention Time	30 to 60 minutes	
pH	11 to 12	9.0
Thickener		
Flow Rate	600 to 3600 gpm	
Retention Time	8.5 to 5.1 hours	
pH	12	< 10.5
Filters		
Drum Speed	0.8 to 1.2 rpm	0.7 rpm
Wash Rate	40 to 80 gpm	60 to 70 gpm
Cake Thickness	1/4 to 3/8 inches	

SECTION 4
PERFORMANCE GUARANTEES AND PERFORMANCE DURING ACCEPTANCE TEST AND DEMONSTRATION PERIOD

4.1 SUMMARY OF PERFORMANCE GUARANTEES

The CEA/ADL dual alkali system was designed to meet seven performance guarantees applying to: SO_2 emissions, particulate matter emissions, lime consumption, soda ash consumption, power consumption, waste solids properties, and system availability. These guarantees were offered LG&E by CEA/ADL and passed through in the contract between EPA and LG&E. A 12-day acceptance test, conducted from July 17 to July 28, 1980, measured the performance of the dual alkali system in terms of the guarantees. The process guarantees were also evaluated throughout the demonstration period to monitor system performance on a long-term basis.

The process guarantees listed below include minor revisions since publication of the project manual containing the preliminary design and cost estimate (EPA-600/7-78-010).

4.1.1 Sulfur Dioxide Emissions

When treating flue gas generated by burning coal containing less than 5% sulfur, the system shall provide such control that emissions from the stack shall be no greater than 200 ppm by volume SO_2 (dry basis), not including SO_2 added from reheater operation, and without additional air dilution. When treating flue gas generated by burning coal containing 5% sulfur or greater, the system shall remove at least 95% of the sulfur dioxide in the inlet flue gas.

4.1.2 Particulate Matter Emissions

In addition to meeting applicable regulations, the system shall also meet Federal New Source Performance Standards for emissions of particulate matter under all conditions of boiler operation. The dual alkali system shall not add any particulate matter to the emissions of particulate matter received by the system from the LG&E Cane Run Unit 6 electrostatic precipitator.

The Federal New Source Performance Standards for particulate matter emissions stipulate that: (1) the maximum two-hour average allowed is 0.1 lb particulate matter per million Btu heat input, and (2) the opacity shall not exceed 20% (excluding water), except that 40% opacity is permissible for not more than 2 minutes in an hour.

The stipulation that the system shall not add particulate matter means that the total particulate mass loading after the dual alkali system shall not be greater than the mass loading before the dual alkali system.

4.1.3 Lime Consumption

The consumption of lime in the system shall not exceed 1.05 moles of available CaO in the lime feed per mole of SO_2 removed from the flue gas. The lime consumption guarantee may be exceeded by 30% during acceptance testing but must be met during the demonstration period.

4.1.4 Sodium Carbonate (Soda Ash) Makeup

Soda ash makeup shall not exceed 0.045 moles of Na_2CO_3 per mole of SO_2 removed from the flue gas provided that the chloride content of the coal burned averages 0.06% or less. If the average chloride content of the coal is above 0.06%, then additional sodium carbonate will be allowed at the rate of 1/2 mole Na_2CO_3 for each mole of chloride (Cl^-) in the flue gas resulting from chloride in excess of 0.06% in the coal. The Seller as part of the guarantee shall perform the necessary research and design to reduce the makeup requirements of Na_2CO_3 from the guarantee point to a level approaching minimal makeup.

4.1.5 Power Consumption

At the peak operating rate (300 MW), the system shall consume a maximum of 1.2% of the power generated by the unit.

4.1.6 Waste Solids Properties

The waste produced by the vacuum filter shall contain a minimum of 55% insoluble solids.

4.1.7 System Availability

The system shall have an availability (as defined by the Edison Electric Institute) of at least 90% for one year. Thus, the system shall be available for operation at least 90% of the calendar year.

The Edison Electric Institute defines operating availability as "the percent of time the unit was available for service, whether operated or not. It is equal to available hours divided by the total hours in the period under consideration expressed as a percentage."

4.2 ACCEPTANCE TEST AND DEMONSTRATION PERIOD RESULTS

The dual alkali system's ability to comply with the performance guarantees was closely monitored during the acceptance test and throughout the demonstration period. SO_2 emissions were continuously measured. Filter cake samples were analyzed periodically to determine lime and soda ash consumption and the insoluble solids content of the filter cake. Power consumption was directly monitored.

4.2.1 Acceptance Test Results

The purpose of the acceptance test was to assess the performance of the dual alkali system in terms of the process guarantees. Table 12 summarizes those guarantees and the corresponding results of the acceptance test. As shown in Table 12, the FGD system did meet all the acceptance test process guarantees, with the exception of the filter cake properties. Table 13 summarizes system performance on a daily basis.

TABLE 12

PERFORMANCE GUARANTEES AND ACCEPTANCE TEST RESULTS

Guarantee	Test Results
SO_2 Emissions Limit	
200 ppm (dry basis) without additional air dilution	141 ppm (dry basis) without additional air dilution (93.9% SO_2 removal)
Calcium Consumption	
1.05 moles available CaO per mole SO_2 removed	1.04 moles available CaO per mole SO_2 removed
Soda Ash Consumption	
0.045 moles Na_2CO_3 per mole SO_2 removed	0.042 moles Na_2CO_3 per mole SO_2 removed
Net Particulate Addition	
No net particulate addition by FGD system	Net particulate removal efficiency averaged 88%
Power Consumption	
System will consume (excluding reheat) not more than 1.2% of power generated at gross peak boiler capacity (300 MW)	System consumed 1.06% of power generated at highest actual boiler operating load (240 MW)
Filter Cake Properties	
Filter cake will contain a minimum of 55 wt % insoluble solids	Filter cake averaged 52.2 wt % insoluble solids
System Availability	
System will have an availability of at least 90% for one year*	Not applicable to acceptance test

*Later modified to require an operability of 90% or greater, coupled with a daily average SO_2 removal efficiency of 90% or greater.

TABLE 13

ACCEPTANCE TEST RESULTS[a]

Acceptance Test Day	Stack SO_2 Emissions[b] (ppm)	SO_2 Removal (%)	Particulate Removal (%)	CaO Consumption[c]	Na_2CO_3 Consumption[d]	Power Consumption[e] (%)	Insoluble Solids (wt %)	System Availability (%)
July 17	130	94.7	85.6	1.082	0.037	1.05	52.65	100.0
July 18	129	95.0	81.7	1.054	0.038	1.08	52.20	100.0
July 19	130	94.6	89.0	1.093	0.045	1.03	52.60	100.0
July 20	152	93.4	88.2	1.096	0.031	1.07	53.72	100.0
July 21	157	93.1	90.0	1.082	0.029	1.08	53.92	100.0
July 22	140	94.5	91.8	1.044	0.070	0.99	51.90	100.0
July 23	132	93.6	85.6	1.029	0.038	1.04	50.70	100.0
July 24	124	94.1	86.0	1.018	0.032	1.07	51.40	99.6
July 25	146	93.8	88.9	0.983	0.068	1.07	51.00	100.0
July 26	171	92.7	86.9	0.995	0.028	1.08	52.43	100.0
July 27	156	93.1	88.0	1.011	0.047	1.08	52.82	100.0
July 28	130	94.0	—	0.990	0.041	1.06	50.58	100.0
Average	141	93.9	—	1.040	0.042	1.06	52.16	100.0

Notes:

a. Over the 12-day test period, the liquid-to-gas (L G) ratio ranged from 7.8 to 9.8 gal/10^3 acf; the active sodium ranged from 0.40 to 0.54 molar.

b. Values shown are 24-hour average SO_2 results.

c. Value shown is available CaO consumption per mole SO_2 absorbed (i.e., 1.139 total moles CaO/mole SO_2 in filter cake x 0.95 moles available CaO/total moles CaO in carbide lime feed = 1.082 moles available CaO consumed per mole SO_2 removed).

d. Moles Na_2CO_3 consumed per mole SO_2 removed.

e. FGD power consumption during peak boiler load.

According to the original agreement and schedule, the acceptance test was to be conducted before the start of the demonstration year. The first attempt to conduct the acceptance test was made on December 3, 1979. However, this test was terminated on December 7, 1979 due to low solids concentration in the carbide lime feed, high pressure drop across the absorber, and a lack of operable spare recycle pumps. In late April after repeated postponement of the acceptance test, a decision was made to proceed with the demonstration year on May 1, 1980. The acceptance test was to be rescheduled for a later date when all Unit 6 and FGD system components could be fully and simultaneously available, and the boiler operated at, or near, full load.

The acceptance test began at 0000 hrs on July 17, 1980. The program ran smoothly for the first 6 days except for the discrepancy in SO_2 emission results between the Lear Siegler in-stack continuous monitor readings and the EPA Method 6 analysis. This discrepancy was traced to a burned out ground wire in the Lear Siegler system, which resulted in recorded SO_2 emission values that were consistently lower than those noted by EPA Method 6 analysis. Based on readings obtained after repair of the ground wire, the Lear Siegler results for the first 6 days were corrected by adding 30 ppm to the hourly and daily averages.

For each test day, the boiler load was raised to its maximum achievable level (240-250 MW) sometime between 0600 and 0800 hours. The load was limited to this level because the FGD system booster fan could handle no more gas volume. Since the boiler was operated with an absorber inlet oxygen concentration of 7.7% (instead of the design of 5.9%), a larger volume of gas was produced at the corresponding generation rate. The system was operating at maximum design gas flowrate (about 1,000,000 acfm), yet only at 80% of design generating capacity. High load was held daily for 10 hours (an acceptance test prerequisite) and then reduced in steps to a low level of about 80 MW for 6-7 hours, generally between 0000 and 0700 hours, in line with electrical energy demand.

The test proceeded smoothly until July 23 when wet coal, resulting from heavy rains during the previous night, blocked the coal feeders to the mills. The boiler load was restricted to 150-160 MW for 2 days. Process stream analysis and stack sampling continued because the FGD system was performing well, but the test program was extended by 2 days to record 10 days of operation at full capacity. By July 25 the wet coal problems subsided, and the program continued to completion at 2400 hrs on July 28.

4.2.2 Demonstration Period Test Results

Table 14 defines monthly performance parameters for FGD system utilization, availability, reliability, and operability. The system performance in terms of these parameters is shown in Table 15. Significant variations from the process guarantees are discussed below. As Table 15 shows, the FGD system met all performance guarantees except those for filter cake insoluble solids and sodium consumption. These two guarantees were not met primarily because of the filter problems discussed in Section 6.

The power consumption guarantee states that at gross peak boiler load the FGD system will consume a maximum of 1.2% of the power generated by the unit. Many of the power consumption figures shown in Table 15 exceed 1.2%. This is expected, since power consumption is not a linear function of boiler load, and the values shown represent periods of operation well below peak load. (Table 12 indicated that FGD power consumption was only 1.06% at 240 MW, 80% of gross peak load.)

During November 1980, the system met the lime consumption guarantee, but did not meet the guarantees for SO_2 removal, availability, soda ash consumption, power consumption, and filter cake

TABLE 14

DEFINITIONS OF PERFORMANCE PARAMETERS[a]

Period Hours (A)	Hours in the period under consideration
Boiler Hours (B)	Hours during the period that the boiler was actually operated
FGD Hours (C)	Hours during the period that the FGD system (one or both modules) was actually operated
Available Hours (D)	Hours during the period that the FGD system was capable of being operated
Called Upon Hours (E)	Hours during the period that the FGD system was demanded (or wanted) for operation
Boiler Utilization ([B/A] x 100)	Hours that the boiler was actually operated divided by the period hours multiplied by 100
FGD Utilization ([C/A] x 100)	Hours that the FGD system was actually operated divided by the period hours multiplied by 100
FGD Availability ([D/A] x 100)	The hours that the FGD system was available for operation divided by the period hours multiplied by 100
FGD Reliability ([C/E] x 100)	The hours that the FGD system was actually operated divided by the called upon hours multiplied by 100
FGD Operability ([C/B] x 100)	The hours that the FGD system was actually operated divided by the boiler hours multiplied by 100

[a]**Source:** Edison Electric Institute.

TABLE 15

SUMMARY OF SYSTEM PERFORMANCE IN TERMS OF PERFORMANCE GUARANTEES[a]

	SO_2 Emissions[b] (ppm)	SO_2 Removal (%)	CaO Consumption[c]	Na_2CO_3 Consumption[d]	Power Consumption (%)	Insoluble Solids (wt %)	System Availability (%)
May 1980	130.0	93.2	1.23	0.079	1.10	55.0	99.9
June 1980	115.5	95.0	1.20	0.085	1.15	47.8	99.9
July 1980	143.0	94.4	1.04	0.042	1.05	52.2	99.9
August 1980	131.0	93.6	—[e]	—[e]	—	—[e]	99.3
September 1980	157.5	92.6	1.15	0.058	1.26	49.7	94.0
October 1980	150.0	89.8	1.04	0.076	1.43	46.8	99.9
November 1980	312.0	82.6	0.89	0.132	1.57	36.9	85.1
December 1980	—[f]	—[f]	—[f]	—[f]	—	—[f]	60.0
January 1981	205.0	91.2	1.04	0.066	1.85	49.1	84.0
February 1981	129.8	93.7	1.06	0.065	1.3	50.3	99.6
March 1981	155.7	92.0	1.04	0.069	1.7	50.5	99.8
April 1981	143.8	93.1	0.93	0.045	1.18	52.1	98.6
May 1981	204.9	88.2	0.95	0.068	—	46.1	96.8
Average	164.9	91.6	1.05	0.071	1.36	48.8	93.2[g]
Process Guarantees	200.0	—	1.05	0.045	1.20	55.0	90.0

Notes:

a. Particulate matter removal efficiency was determined during acceptance test only (88%).
b. All values shown (except May 1980) are 24-hour average stack SO_2 emissions, corrected for moisture and reheater. May 1980 results represent the average of the 2 absorbers' outlet emissions.
c. Total moles available CaO consumed per mole SO_2 removed.
d. Moles Na_2CO_3 consumed per mole SO_2 removed.
e. No filter cake samples were analyzed in August 1980.
f. FGD system operated only 14.4 hours in December. No system characterization tests were performed.
g. Average availability based on hours in total period (see Table 14).

composition. The poor performance was largely attributed to the presence of significant amounts of suspended solids in the thickener overflow. Extended periods of operation at low boiler load and high oxidation rates caused the ratio of sulfate:sulfite in the scrubbing liquor to exceed 4:1. At this level of sulfate, the settling characteristics of FGD solids deteriorated, and a significant volume of FGD solids was carried over the thickener weirs and back into what should have been the clear liquor section of the process. Solids carryover from the thickener overflow increased the rate of absorber plugging and reduced the SO_2 removal capability. Lime consumption was extremely low, but this was at the expense of high consumption of soda ash, a more costly material. The high soda ash consumption resulted from reduced washing of the filter cake.

In December 1980, the FGD system operated only for 14 hours of the 240 hours that the No. 6 boiler was in operation, and none of the performance guarantees were met. The boiler was shut down primarily for scheduled maintenance, and during this outage the FGD system was considered available. However, after the boiler had come back on-line, a thickener rake problem rendered the FGD system inoperable for about 300 hours. The rake in the thickener had separated from the drive shaft. The FGD system and the thickener were briefly operated (14 hours) to keep the thickener slurry from freezing and were then shut down so the rake could be repaired. No system characterization tests were performed that month.

SECTION 5
SUMMARY OF MONTHLY PERFORMANCE PARAMETERS

The operation of the FGD system was monitored monthly on the basis of several performance parameters. These parameters, as defined by the Edison Electric Institute, were shown in Table 14. The FGD system performance in terms of these parameters is summarized in Tables 16 (by month) and 17 (demonstration period).

Major FGD system outages for the period shown in Table 17 were due primarily to difficulties with the thickener and boiler outages. The significant operating problems in December 1980 were caused by excessive solids accumulation in the thickener and separation of the thickener rake from the drive shaft. Thickener problems were again responsible for the FGD system downtime in June 1981, when solids were allowed to accumulate in the thickener because of a false reading indicating low rake torque. Boiler outages accounted for most of the FGD system downtime in September 1980. Section 6 discusses in detail the extent to which FGD availability, reliability, and operability were a function of FGD system operating problems.

TABLE 16

SUMMARY OF MONTHLY PERFORMANCE PARAMETERS FOR FGD SYSTEM

	Utilization	Availability	Reliability	Operability
May 1980	99.5	99.9	99.9	99.9
June 1980	88.9	99.9	99.9	99.7
July 1980	83.1	99.9	99.9	99.8
August 1980	84.5	99.3	94.6	94.6
September 1980	86.9	94.0	98.8	98.6
October 1980	99.9	99.9	99.9	99.9
November 1980	55.1	85.1	96.8	96.8
December 1980	1.9	60.0	6.0	6.0
January 1981	80.2	84.0	83.3	83.3
February 1981	71.6	99.6	99.5	99.5
March 1981	71.9	99.8	99.7	99.7
April 1981	82.6	98.6	98.3	98.3
May 1981	96.8	96.8	96.8	96.8
Demonstration Period*	77.4	93.2	94.1	94.0

*Demonstration period averages were calculated as shown in Table 17.

TABLE 17

AVERAGE PERFORMANCE PARAMETERS FOR DEMONSTRATION PERIOD (May 1, 1980 — MAY 31, 1981)

Period Hours (A)	9048
Boiler Hours (B)	7451.5
Percent Boiler Utilization (B/A)	82.36
FGD Hours (C)	7005.2
Hours FGD Available (D)	8435.2
Hours FGD Called Upon (E)	7447.3
Percent FGD Utilization (C/A)	77.42
Percent FGD Availability (D/A)	93.23
Percent FGD Reliability (C/E)	94.06
Percent FGD Operability (C/B)	94.01

SECTION 6
START-UP AND DEMONSTRATION PERIOD OPERATING PROBLEMS AND RESOLUTIONS

Several problems were encountered during the start-up and operation of the LG&E Cane Run dual alkali system. Most of these problems can be classified as being either instrument, mechanical equipment or process related, but some had more than one cause.

6.1 PROCESS-RELATED OPERATING PROBLEMS

Process-related operating problems included the following:

- Winter fill and start-up of dual alkali system.
- Carbide lime (solids concentration, alumina content, and reactivity).
- Boiler (load swing, air leakage, and thickener settling, as explained in Section 6.1.3).
- Fly ash carryover (ESP operation).
- Filter operation and sodium recovery.

6.1.1 Winter Fill and Start-up

Before the system was initially filled with sodium carbonate (soda ash) solution, each of the individual process sections was examined. The instruments were calibrated, and the operation of all mechanical components was checked. After several mechanical problems had been corrected (such as reactor agitator overload and reactor flow short-circuiting), filling of the system was begun.

In normal operation, makeup soda ash is dissolved in thickener return water (which has a temperature in excess of 100°F) and then returned to the scrubber. During the initial filling of the dual alkali system in March 1979, however, the soda ash did not dissolve easily because of the low temperature (50-54°F) of the fresh makeup water supplied to the soda ash tank. A small package boiler was required to preheat the water before it was pumped to the soda ash tank.

After the thickener and thickener (overflow) hold tank had been filled with (warm) soda ash solution, soda ash solution was pumped to each of the absorber modules. When the absorber modules were filled to operating level, liquor was bled to the reactor train until it also was filled to overflow. At that time, the flow loop back to the thickener was complete. The absorbers were operated intermittently to warm the scrubbing liquor while liquor from the thickener (hold tank) was recycled through the soda ash solution tank back to the thickener with continued soda ash addition. When the active sodium concentration in the scrubbing system reached 0.35 M, lime feed to the reactor trains was started, and the absorbers were placed in operation.

6.1.2 Carbide Lime

6.1.2.1 Carbide Lime Solids Concentration — Obtaining carbide lime slurry with solids concentration above 20% was a problem during the system start-up. [This continued until modifications were made in the carbide lime barge loading system at Airco (the carbide lime supplier) and the lime grinding/classifying system at LG&E in February and March 1980.] Excess water introduced with the lime at low solids concentration upset the water balance of the scrubbing system. The water introduced with lime feed and filter cake wash exceeded the water losses in the filter cake and scrubber evaporation. The resulting increased water inventory eventually filled the thickener to the overflow level. At this point, it was necessary to discontinue the lime feed until enough water had been evaporated from the scrubbing liquor by the hot flue gases in the scrubber to return the thickener level to normal. Unfortunately, when the lime feed was discontinued, the process scrubbing liquor became acidic, and SO_2 removal efficiency dropped below the required 90%.

A temporary solution was to decant the dilute lime slurry (from the barges) in the lime storage tanks and pump the settled slurry to the day-tank. Because lime requirements for the scrubbers could not always be coordinated with barge unloading schedules, high-solids lime could not always be supplied to Unit 6. Periodic water balance problems persisted until Airco completed its larger pipeline to deliver concentrated lime to the transport barge and the LG&E carbide lime grinding and classifying system became fully operational.

6.1.2.2 Alumina Dissolution and Tray Scaling — Formation of scale deposits on the absorber trays had been evident from start-up. The trays were recleaned whenever an outage of the boiler or scrubbing system permitted. When scrubber operation resumed, so did scale formation. Eventually enough of the tray holes would become blocked and pressure drop across the scrubber would become high enough to require a shutdown for tray cleaning.

ADL and Bechtel analyses of the tray deposits showed that the scale was mainly a complex of alumina and silica (Table 18). Subsequent testing by ADL demonstrated that alumina in the carbide lime dissolved at a pH of 11.4 and precipitated when the pH dropped below 10.7.

The original design and operating conditions for the system called for pH's of:

- 6.0 for the absorber,
- 11.5 for the secondary reactor,
- 12 for the thickener overflow, and
- 7 to 10.5 for combined thickener overflow and absorber recycle stream.

Unfortunately, these conditions were ideal for dissolving aluminum in the reactors and thickener and then precipitating an alumina-silica complex on the absorber trays. Table 18 presented selected analyses of scale taken from the absorber trays. Figure 5 shows how laboratory batch reactor tests were conducted to determine the effect of pH on aluminum and silica concentrations in the scrubbing liquor (and the potential for scaling). As carbide lime is introduced to the reactor system, the increase in pH causes aluminum and silica (present in the lime) to become dissolved in the scrubbing liquor. The pH subsequently drops when the solution reacts with sulfur dioxide in the scrubber, causing the alumina-silica complex to precipitate out of solution. After this mechanism of scale formation had been established by laboratory testing (Figure 5), process solutions from the full-scale system were analyzed and showed the same relationship between pH and concentrations of aluminum and silica in the scrubbing liquor (Table 19).

TABLE 18

ANALYSIS OF ABSORBER TRAY SCALE DEPOSITS

Sample No.	Sample Location (Absorber Module)	Sample Date	Analysis (%)						Laboratory Performing Analysis
			TS (as SO_4)	Calcium	Iron	Aluminum	Silica (SiO_2)	Acid Insol.	
497	(A) "berry"*	4/27/80	4.6	0.50	4.5	9.74	4.15	48.8	Bechtel
498	(A) upper tray	4/27/80	14.0	0.20	0.61	18.92	25.95	3.9	Bechtel
499	(B) upper tray	4/27/80	11.0	<0.10	0.31	19.88	21.65	10.4	Bechtel
500	(A) lower tray	4/27/80	15.5	0.30	0.45	17.55	24.05	3.9	Bechtel
—	(A) "berry"*	12/79	—	0.36	8.1	12.8	5.6	27.0	ADL

*A berry-shaped stalagtite that formed around the bottom of the holes in the absorber trays.

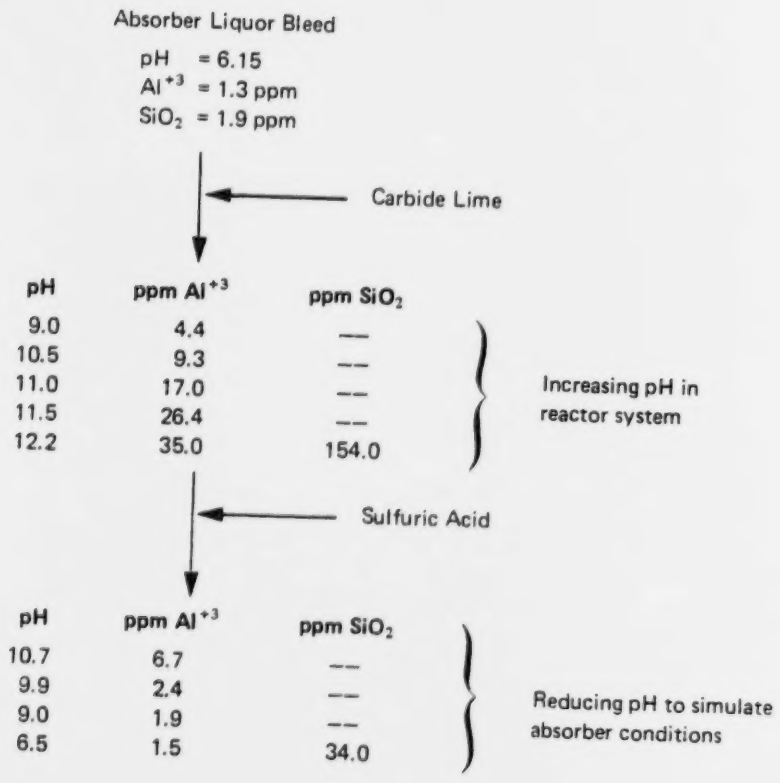

FIGURE 5 LABORATORY BATCH REACTOR TESTS: ALUMINUM AND SILICA DISSOLUTION FROM CARBIDE LIME

TABLE 19

ALUMINUM AND SILICA CONCENTRATIONS MEASURED IN CANE RUN UNIT 6 SCRUBBER PROCESS SOLUTIONS

Date		Lime Slurry	Reactor 1	Reactor 2	Thickener Underflow	Thickener Overflow	Absorber Bleed
10/5/79[a]	pH	—	>12	>12	—	>12	5.6
	$[Al^{+++}]$	<1 ppm	32 ppm	40 ppm	—	35 ppm	22 ppm
4/16/79[b]	pH	—	—	9.45	11.75	11.25	6.1
	$[Al^{+++}]$	<1 ppm	—	2.4 ppm	21.9 ppm	11.8 ppm	10.8 ppm
4/17/79[b]	pH	—	—	9.0	11.25	9.75	5.9
	$[Al^{+++}]$	—	—	2.8 ppm	12.6 ppm	2.3 ppm	4.1 ppm

a. Analyses by Arthur D. Little, Inc.
b. Analyses by Louisville Testing Laboratories

Accordingly, the operating pH conditions were changed from those originally specified to:

- 6.3 for the absorber,
- 8.0 for the primary reactor,
- 9.0 for the secondary reactor, and
- <10.5 for the thickener overflow.

The absorber liquor pH was raised to compensate for the lower pH of the thickener overflow fed to the top tray (and to maintain SO_2 removal efficiency). This was accomplished by increasing the "feed forward" flow of liquor from the thickener to the absorber.

Incorrect measurement of absorber recycle liquor and thickener hold tank liquor flow to the trays is believed to have resulted in dry spots on the trays during operation and added to the severity of the tray scaling problem, as explained in Section 6.2.2.

6.1.2.3 Carbide Lime Reactivity — Before the lime grinding and classifying system at Cane Run Unit 6 became operational, carbide lime was settled in the "B" lime storage tank (to increase solids concentration) and then run through an in-line comminutor and coarse screen to the day storage tank. Although this screened lime was handleable in the dual alkali system, the coarse inerts had a tendency to act as nuclei for the formation of sulfate/sulfite nodules in the reactors, and reactivity of the carbide lime (moles of lime required to remove a mole of SO_2) was below guarantee.

When the grinding/classifying system was installed, hydroclones were used to separate out the coarse fraction, which was sent to the ball mill system. From there, the ground coarse fraction was routed to lime storage tank "A" for other Cane Run scrubbing systems. The fines went to the "B" lime storage tank for use in Unit 6. With the grinding/separation system in operation, the incidence of lime overfeeding and nodule formation decreased significantly, and lime consumption dropped to 1.05 moles "available" calcium consumed per mole SO_2 removed, meeting the performance guarantee. The term "available" calcium is defined as the portion of the total calcium available (as the oxide or hydroxide) at a pH of 10.2, the pH of the secondary reactor. This correction term for total calcium is necessary to compensate for non-reactive calcium (such as carbonate) and other inerts in the carbide lime. (For specifications of the carbide lime, see Table 2, Section 3.)

6.1.3 Boiler

6.1.3.1 Daily Load Swing — To limit slagging of the boiler tubes, Unit 6 must operate at reduced load overnight, usually about 80 MW. Each morning, the boiler is brought up to an operating load that varies depending on power demand and boiler condition.

Although the dual alkali system was designed to handle a boiler load ranging from 60-300 MW, the system will not operate with optimum efficiency under continually changing operating conditions. Such variations affect SO_2 removal efficiency, reactor train pH control, and thickener settling. Shortly after these conditions stabilize each day, it is time again for the evening boiler load reduction. (See Section 6.2.4 for an explanation of the effect of boiler load change on reactor train pH control.)

6.1.3.2 Air Leakage — The oxygen content of the flue gas is the result of excess air for combustion, in addition to air leakage in the boiler, preheater and electrostatic precipitator (ESP). The scrubbing system was designed for an oxygen content of 5.7 (vol. %). Actual oxygen concentrations in the flue gas have

exceeded 8%. This has had two effects. First, the scrubbing system reaches its maximum operating pressure drop at a boiler gross output of about 260 MW rather than at 280-300 MW as designed. Second, the higher level of oxygen leads to higher oxidation rates of the scrubbing liquor and impaired thickener performance, as discussed in Section 6.1.3.3.

6.1.3.3 Thickener Settling — For a while during boiler operation at low maximum load (80-140 MW), both absorbers were operated 24 hours a day. Low SO_2 input together with a high oxidation rate (at low boiler load, the % volume of excess air is greater) caused the sulfate/sulfite ratio in the scrubbing liquor to increase to >4:1. At this level of sulfate, the settling characteristics of FGD solids deteriorated, and a significant volume of unsettled fine solids was carried over the thickener weirs and back into what should have been the clear liquor section of the process (hold tank and absorbers).

To reduce solids carryover from the thickener, the number of absorber modules in operation was reduced from 2 to 1 for boiler loads below 120 MW. This decreased the number of trays contacting flue gas (oxygen) with liquor from four (with two absorbers) to two trays (with one absorber) and seems to have prevented recurrence of the high sulfate levels observed when the thickener solids carryover was most severe.

Other thickener problems have included inaccurate rake torque measurement and solids accumulation during shutdown of the CSI waste disposal system.

6.1.4 Fly Ash Carryover (ESP Operation)

The Unit 6 dual alkali scrubbing system was designed for an inlet (fly ash) particulate loading of <0.1 $lb/10^6$ Btu. Fly ash particulate loading to the scrubber has averaged about 0.85 $lb/10^6$ Btu, over eight times the design figure. During upsets in ESP operation the loading has often been significantly above this average.

Although the scrubbers were not designed for particulate removal, they have effectively removed a high percentage of the fly ash from the flue gas. The scrubber modules, designed for clear liquor operation, however, do not have agitators. As a result, fly ash solids have accumulated in the bottom of each module. In January 1980, during a shutdown of the scrubbing system for work on the by-pass damper, both absorbers were cleaned of accumulated fly ash solids, and the solids were flushed to the thickener. The thickener rake jammed because of the solids overload. The thickener was drained, a hole was cut in its side, and the solids were removed with a front-end loader.

In addition to the solids accumulation problem, the fly ash also contributes dissolved aluminum (possibly with the potential for scaling) at a pH of over 11.0.

While excessive fly ash carryover due to ESP malfunction caused pluggage of the scrubber lines and valves and increased the load on the thickener, the scrubbing system does seem to tolerate the average Unit 6 ESP outlet fly ash loading. After a stabilized bed of fly ash solids forms on the scrubber module floor, excess fly ash is gradually carried over to the thickener without impairing thickener operation.

6.1.5 Filter Operation and Sodium Recovery

Sodium losses (and makeup soda ash requirements) have averaged about twice the guarantee of 0.045 moles soda ash per mole SO_2 removed. It has been possible to obtain the proper filter cake thickness

and wash rate under controlled tests and reduce sodium losses to less than 1/2 the guarantee, but under long-term operation these rigid filter conditions have not been held.

Balancing thickener sludge solids concentration, filter drum speed, filter tub slurry depth, and filter feed rate to give consistent cake thickness (~3/8") has been a continuing problem, and the proper wash rate is rarely maintained for various reasons. (These include operator lack of appreciation for importance of wash rate and water imbalance situations caused by excessive makeup water addition.) The high sodium losses will likely continue unless a more highly trained crew is used to operate the filters or the rotary vacuum filters themselves are replaced with horizontal belt filters of the type used in the other two commercial utility dual alkali FGD systems. The proper filter operating conditions were listed in Table 8.

6.2 INSTRUMENT-RELATED OPERATING PROBLEMS

Inaccurate pH measurement was the most frequent instrument-related operating problem. Operating the scrubber modules at too high a pH caused excessive liquor flows in the system and reduced regeneration reactor and thickener settling time. Operation at low pH, on the other hand, reduced SO_2 removal efficiency. Inaccurate pH measurement of the scrubbing liquor in the reactor train usually led to lime overfeeding. This, in turn, would cause dissolution of aluminum (and subsequent scaling of the absorber trays) or formation of nodules that collected in the bottom of the reactors.

In the manual operating mode, the liquor flow must be set manually. Drift in calibration of the scrubbing liquor flowmeters may also have contributed to tray scaling by indicating adequate liquor flow to the trays when, in fact, the flow may have been low enough to create dry spots on the trays.

Another instrument problem was evidenced by discrepancies in absorber outlet SO_2 measurements. These discrepancies initially made it difficult to balance the operation of the two absorbers and to reconcile the absorber outlet SO_2 measurements with the stack measurements.

6.2.1 pH Probes

Decreased sensitivity of the pH probes resulted primarily from scale buildup on the probes. A thin, almost transparent film of calcium sulfate (gypsum) would accumulate on the absorber bleed pH probes over a period of days until the sensitivity was reduced to a point where the absorber could no longer be run on automatic pH control. The absorber would then be operated in manual mode until the pH probes could be removed, cleaned, recalibrated and replaced. (In the manual control mode, the flow of thickener overflow to the absorber is set manually at a constant rate depending on boiler load and stack SO_2 concentration.)

A heavy white scale of calcium sulfate also formed on the probes in the secondary reactors and would typically become thick enough in 24 hours to significantly reduce response time (minutes versus seconds response). With a delayed response time, the potential for periodic overfeeding of carbide lime was greatly increased, since lime feed would continue until the probe sensed the effect (increase in pH) of the lime addition. Lime overfeeding then resulted in carbide lime alumina dissolution (and absorber tray scaling) or nodule formation in the reactors.

LG&E recognized the need for continuing pH probe calibration and implemented a program to check pH probes daily. The probes are now acid-cleaned and calibrated as required. Initially, the probes were cleaned with hydrochloric acid, but this required a very long soak period and sometimes left a thin film of deposit. LG&E laboratory personnel tried various other cleaning agents and found that dilute nitric acid worked rapidly to completely remove the deposit.

The frequency with which the pH probes must be cleaned to maintain the necessary response time has not yet been finalized, but it appears that the absorber bleed probes should be cleaned every other day. The secondary reactor probes must be cleaned daily to maintain an adequate response time for automatic control.

Originally, the absorber pH probes were installed in pressure lines running from the recycle pumps to the absorber modules. Repeated removal of these probes for calibration checks and cleanings required a great deal of time. In addition, the flexing and twisting of the pH probe lead wires during each removal led to their rapid failure. To remedy this situation, LG&E placed the absorber bleed pH probes in holders in a small stainless open-top trough from which they could be easily removed for cleaning and recalibration. The trough was fed from a line from the recycle pump and overflowed to the floor drain.

Similarly, the pH probes in the secondary reactors, originally hardwired into a conduit at a depth of about 20 feet, were replaced with flexible pH probes wired into a lighter weight probe holder about 12 feet below the surface of the reactor liquid. This arrangement allowed the probes to be more easily serviced.

Placing pH probes in the discharge chute from the primary reactor to the secondary reactor was also tried as an alternative means for pH control of the reactor train. This location would have allowed a much more rapid response to pH change in the reactor system. Unfortunately, the pH probes scaled even more rapidly in the discharge chute than they had in the secondary reactor. The pH control probes were returned to the secondary reactors.

6.2.2 Flowmeters

In-line magnetic flowmeters monitored the flow rate of thickener overflow and absorber tray recycle. The flowmeters were flow-calibrated when the system was initially started up (before the demonstration period) and then periodically recalibrated electronically. A cross-check of flow rates with a clamp-on type doppler flowmeter about two months before completion of the demonstration period indicated that, in some cases, the actual flow rates were less than 50% of those indicated by the magnetic flowmeters.

Since the actual thickener hold tank and absorber recycle flow rates to the trays were much less than indicated by the flowmeters, it is very likely that portions of the trays were frequently not covered with liquor and operated dry. Indeed, during several inspections of the trays to monitor for any scaling, little or no scaling was found, but some tray holes were plugged with dried (soluble) salts. The reduced liquor flow may not have been directly responsible for the insoluble aluminum salt scale that often plugged the trays. However, operating the absorber with dry spots on the trays may have contributed to the deposition of this aluminum salt scale. (Section 6.1.2.2 discusses the source of these aluminum salts.)

6.2.3 SO_2 Removal Efficiency Monitoring

The primary method of measuring SO_2 removal relied on the continuous Lear Siegler monitor (SM810) installed in the stack and the du Pont Model 460 analyzer (which alternately samples the two scrubber inlet gas streams). Inlet flue gas O_2 concentrations were monitored by *in situ* Hagen O_2 analyzers, while stack gas O_2 concentrations were continuously monitored by a Lear Siegler monitor (CM-50).

Shortly after system start-up, a difference between the SO_2 removal efficiencies of Modules A and B (as measured by the du Pont monitors in the scrubber inlet and outlet) was noted, even though both absorbers were operating at the same flow rates and pH conditions. This difference in SO_2 removal

efficiency was not affected by recalibration of the monitors. The connections between the outlet SO_2 probes and du Pont analyzers of Modules A and B were switched and results indicated that the outlet SO_2 probe location was the source of the discrepancy. LG&E subsequently tested the gas velocity and SO_2 concentration profiles across the scrubber outlet ducts and found that a stratified gas flow was causing the differing SO_2 concentrations. The SO_2 probes were relocated further downstream in the scrubber outlet ducts.

Wet chemical analyses of the stack gases, performed during the acceptance test, showed discrepancies in the Lear Siegler stack monitor readings. An extensive check of the system revealed a burned-out ground wire in the signal line of the Lear Siegler continuous SO_2 monitor. On the basis of analyzer calibration data before, during, and after elimination of the grounding problem, it was concluded that the damaged ground wire had caused analyzer signals that were 30 ppm lower than reality. After the ground wire was repaired, results from the wet tests correlated well with the data from the Lear Siegler monitor.

6.2.4 Reactor pH Control

Control of pH in the reactor train (and thickener) is critical to the continuous operation of the Cane Run dual alkali system. As described more fully in Section 6.1.2.2, alumina dissolution from the carbide lime occurred when reactor train/thickener systems were operated at high pH levels. Subsequent precipitation of this aluminum as insoluble aluminum salts in the absorber caused severe scaling and tray plugging.

In the original design of the reactor pH control system, lime addition to the primary reactor was directly controlled by the pH measured in the secondary reactor. A lime feed control override was also incorporated to prevent lime overfeeding during rapid reductions in absorber bleed rates typically encountered during a reduction in boiler load.

Unfortunately, lime deposits frequently plugged the lime feed valves and would subsequently break free, causing wide swings in the pH of both the primary and secondary reactors. The system was therefore modified to maintain a relatively constant lime addition rate and minimize variations in pH by using the lime slurry flowmeters in the lines to the primary reactors to control the lime feed valves at a set flow rate. The lime feed rate to each of the primary reactors was manually set by the operators in response to the pH readings in the secondary reactors as read on the control room board. The frequency with which the thickener pH then exceeded 10.5 (the pH at which aluminum dissolution and tray scaling occurs) depended on the skill of the individual operators.

To allow automatic control of pH and limit the swing of pH in the secondary reactors to a narrower range than could be maintained manually (by the operators), the reactor pH control system was further modified, and secondary reactor pH is now used to automatically control the setpoint for lime flow to the primary reactors. The flow rate set by the pH controller is held constant by feedback from the lime flowmeters to the lime feed control valves. This has worked very well at constant boiler load (nearly constant lime demand) but not so well as the boiler undergoes its normal daily cycle from 80 to 240 MW. When the boiler load and scrubbing liquor bleed rate change, the amplitude of the pH oscillation about the control point increases and then narrows again after the boiler load has stabilized.

One further modification in the reactor pH control system is contemplated. The absorber bleed flow rate will be used to set the lime feed rate control point, with the pH controller used to trim the lime feed rate setting. This should reduce the wide oscillation in reactor train pH that occurs with change in boiler load and hold the secondary reactor pH within the desired operating range (9-10).

6.3 MECHANICAL EQUIPMENT-RELATED OPERATING PROBLEMS

Various mechanical equipment problems (and lack of high solids carbide lime) delayed the start of the demonstration period until approximately one year from initial startup. The initial startup of the system was in April 1979, and the demonstration period ran from May 1, 1980 to May 31, 1981. While the collapse of the mist eliminator and the thickener plugging caused the longest periods of downtime, other recurrent problems such as piping failure and filter repair continue to add to total downtime and maintenance cost. These and other mechanical problems are discussed below.

6.3.1 Chevron Mist Eliminators

After several weeks of initial scrubber operation, an increase in pressure drop across both the scrubber modules was noted. The scrubbers were inspected and it was discovered that in addition to some deposit formation on the trays, the chevron-type Noryl® mist eliminators had deformed, closing the gas passage. An attempt was made to repair the mist eliminators on site, but this approach was finally abandoned, and a new set of chevron mist eliminators was ordered from a different supplier. These were also fabricated from General Electric's Noryl® plastic, but the new chevrons had considerably more support bracing. After these new Noryl® mist eliminators failed again (in 1982), they were replaced with fiberglass-reinforced-plastic (FRP) mist eliminators of a type that has performed well in other scrubbers at Cane Run.

6.3.2 Thickener

There have been several problems associated with the thickener. Some of these have been process-related, such as high sulfate concentrations or high liquor rates through the thickener. These process-related problems are discussed in Section 6.1.3.3. The equipment-related problems have included incorrect installation of the thickener center well and failure of the rake torque measuring system.

Due to an error in the equipment drawings, the thickener center well was installed 30 inches lower than design depth. As a result, some of the feed to the center well was short-circuiting in the thickener by overflowing the center well and passing directly to the overflow weir. The flow through the center well was also impinging on the sludge blanket in the bottom of the thickener, resuspending settled solids. Reinstallation of the center well at the correct height helped reduce the solids carryover from the thickener.

More recently, it has been discovered that the measurement of thickener rake (hydraulic) torque has been incorrect. This has led to solids overfill, with some solids being carried over the thickener weirs. The resulting high solids inventory in the thickener may also have been the reason the thickener rake stalled and sheared off in December 1980. The metal expansion bellows used to measure hydraulic pressure (and thickener rake torque) had been leaking, leading to low readings of rake torque and solids level in the thickener, when in actuality, the rake torque and solids level were high. Repair of this torque-measuring device should greatly reduce the incidence of excessive solids buildup in the thickener and subsequent thickener rake failure.

6.3.3 Scrubbing Liquor Regeneration Reactors

When lime feed to the regeneration reactors was initiated during startup, the current drawn by the agitators in both the primary and secondary reactors exceeded safe limits for the motor drives. Examination of the design criteria revealed that the motor and drives were undersized for the agitator design. Since replacing both motors and drives would have been costly and would have delayed startup, the impellers were modified to reduce power consumption.

It was then discovered that, because of the direction of agitator rotation and the location of the overflow pipe in the secondary reactors, most of the flow from the primary reactors was passing directly to the overflow of the secondary reactor. This short circuiting was eliminated by installing a 16-foot chute (down-pipe) fastened to the reactor baffles. This chute delivered input flow from the primary reactor to the area near the secondary reactor impeller and provided much more efficient mixing within the tank.

The secondary reactors were designed to operate at varied liquid levels to allow varied retention times. Level controllers, discharge pumps, and discharge valves were to be used to control reactor liquid level. However, it was found that the pressure developed by the discharge pumps was more than the discharge valve could handle without violent vibration at low flow. The secondary reactors were, therefore, operated henceforth on overflow rather than level control.

Maximum retention time was obtained by operating the secondary reactors on overflow to the thickener, and it was subsequently found that this maximum retention time was required to complete the reaction between the carbide lime and scrubbing liquor.

6.3.4 Lime Feed Valves

Pneumatically-operated pinch valves were initially used to control lime feed to the primary reactors. Unfortunately these valves not only had a tendency to wear out rapidly through erosion, but also did not control lime feed accurately enough at low flows. LG&E replaced the pinch valves with teflon-lined stainless steel plug valves of a type that had worked well in their other Cane Run lime scrubbing systems. These replacement valves have worked quite well in the dual alkali system.

Proper location of valves is important in any slurry handling system. A horizontal stretch of pipe is usually best if the line cannot be completely drained when flow is stopped. The lime slurry feed valves at Cane Run were originally located in a section of vertical pipe. When the slurry feed was turned off, the solids often settled in the vertical pipe section above the valve inlet and plugged the line to the valve. Had the valve been located in a horizontal section of piping instead, the settled solids would not have entirely blocked off the valve inlet and would have resuspended once flow was resumed. The lime feed valves on the Unit 6 dual alkali system have been relocated to a short, horizontal section of pipe close to the lime feed circulation loop.

6.3.5 Vacuum Filters

The settled solids from the thickener are filtered and washed to recover sodium salts before discharge to the CSI sludge stabilization system. Obtaining the proper cake thickness and wash efficiency has proven to be a difficult process problem (see Section 6.1.5), and the filters have required substantial maintenance.

The filters installed on the dual alkali system were designed for blowback and scraper operation. Because of a high incidence of torn filter cloth, however, the scraper blades have been removed and cake discharge is now by cloth blowback only. Cloth blowback operation greatly extends filter cloth life (from weeks to months), but filter cake removal efficiency is reduced to 70-80%, with 20-30% of the cake returning to the filter tub.

There has been a continuing program to evaluate different types of filter cloths to find a cloth with the best combination of durability, resistance to blinding (pluggage), and moisture removal efficiency. The cloth installed on the filters originally was a polypropylene monofilament. This cloth, of a relatively open weave, operated at low vacuum and did not produce a sufficiently dry filter cake. For the acceptance test, the polypropylene monofilament was replaced with a nylon multifilament (#7144) cloth with a lower permeability (10-15 CFM). This new filter cloth did operate at a high vacuum and produce a drier cake (still not up to guarantee, however), but had a tendency to severely blind after about 3 to 4 weeks of operation. Two of the three vacuum pumps were also damaged by operation at excessive vacuum (25 in. Hg vs 22 in. Hg design vacuum) during the acceptance test. A more permeable (20 CFM) nylon multifilament (#1111) was then tested. This cloth had more resistance to blinding but did not produce a filter cake consistently dry enough for CSI processing. Finally, a nylon multifilament cloth (#1143) with a permeability between those of the #1111 and the #7144 cloths (10-20 CFM) was tested. This cloth appears to offer the best compromise between drying efficiency (cake dry enough for CSI processing) and resistance to blinding (better than one month's use without serious blinding).

6.3.6 Pumps

Two major changes in the Worthington scrubber recycle and thickener overflow return pumps have been made since scrubber start-up. These were: (1) replacement of the two-piece welded impellers with single-piece cast impellers and (2) replacement of the two-piece suction-side pump liner with a bonded liner. The welded impeller was replaced because of a high incidence of weld failure (shearing between the hub and impeller). Leakage of the two-piece, unbonded pump liner had corroded the bolts holding the liner in place, and this caused the liner to be drawn into the impeller.

6.3.7 Fiberglass Piping

There were several serious failures of the cast fiberglass piping and fittings. Most breaks occurred in areas experiencing vibration or mechanical stress — the absorber recycle lines, thickener overflow return and underflow lines. In some critical areas, such as the thickener underflow lines in the tunnel (where a break flooded the tunnel with three feet of slurry) the fiberglass piping was replaced with carbon steel.

6.3.8 By-Pass Damper Leakage

Gas leakage through the scrubber by-pass damper has consistently exceeded specification (1%). This has periodically caused scrubbed gas to be recirculated through the system and unscrubbed gas to be by-passed to the stack. Linkages to the damper louvers have been frequently adjusted. An additional set of actuators and limit switches, installed in January 1980, has reduced by-pass damper leakage, but not down to the 1% leakage specified.

6.3.9 Soda Ash Feeder

Several problems were encountered with the dry soda ash feed. The most important involved moisture entering the silo, particularly in damp rainy weather. As a result, hard lumps or "rocks" of soda ash formed and jammed the feeder. Numerous attempts were made to seal the silo against moisture, but there was still some leakage and hard lumps of soda ash occasionally jammed the feeder.

To solve another problem, a small blower was installed in the chute between the feeder and the dissolving tank. This was to prevent moisture (from the dissolving tank) from caking the soda ash on the side of the chute and blocking it.

6.3.10 CSI System

It is necessary to suspend operation of the dual alkali vacuum filters for 8-12 hours each day while the pug mill of the CSI sludge fixation system is cleaned. In most cases this has not interfered with operation of the FGD system, since an adequate volume of solids is usually removed from the thickener prior to shut down of the filters. In several instances, however, this has meant that the filters have not been operational when the thickener was filled to capacity with solids and may well have contributed to incidents where solids overflowed the thickener weirs. This situation was accentuated by the problem with failure of the rake torque sensor. Thickener rake torque was used to measure solids inventory in the thickener and determine when to operate the filters. With this device measuring incorrectly, there were probably many more instances where the filters should have been placed in operation but were not.

SECTION 7
COMMERCIAL LIME TEST

7.1 OBJECTIVES

Airco carbide lime, a byproduct of acetylene production, is normally used to regenerate the sodium-based scrubbing liquor at Cane Run. A month-long test with commercial-grade lime in place of carbide lime was conducted as part of the demonstration to compare the performance of the two materials in a lime-based dual alkali system. The specific goals of the commercial-grade lime test were to:

- Operate the reactor at a higher pH to allow lower liquor flow rates and longer thickener settling time.
- Increase the rate and control of the regeneration reaction.
- Improve the settling and filtering properties of the waste solids.

The overall conclusion from the commercial-grade lime tests was that the FGD system performance with commercial lime did not differ significantly from its performance with carbide lime. The anticipated improvements in solids settling and filtration were not realized.

The commercial (dry) hydrated lime, obtained from Mississippi Lime Company, was air-conveyed from the delivery trucks through a jet eductor into the "B" lime storage tank. It was then pumped to the dual alkali day lime storage tank. Table 20 compares the typical specifications for the Mississippi Lime Company with the specifications for Airco carbide lime.

Because the commercial lime test was to be performed at high reactor pH (>11.0), there was an increased potential for aluminum dissolution (from fly ash carried over to the scrubber) and subsequent scaling of the scrubber trays. Aluminum compounds are normally introduced to the system by both the fly ash and carbide lime. At reactor pH above 10.0, the aluminum dissolves in the scrubbing liquor. A scaling problem then results in the scrubber where, at reduced pH (around 6.5), aluminum in the regenerated liquor precipitates out of solution as a silica-alumina complex. To alleviate this problem, the normal operating pH set point of the secondary reactors has been lowered to 9.0 from the original design condition of 11.5 for carbide lime.

Since one of the goals of the commercial lime test was to determine the effects of increasing the thickener pH to more than 11.0, precautions were taken at the start of the test to reduce the amount of aluminum present in the system and thus minimize the potential for aluminum scale. (The aluminum dissolution problem is discussed in Section 6.1.2.2.) The carbide lime was pumped to the scrubbers and to a second storage tank in order to empty the "B" large lime storage tank for the commercial lime. In addition,

TABLE 20

COMPARISON OF MISSISSIPPI LIME COMPANY COMMERCIAL HYDRATED LIME AND AIRCO CARBIDE LIME

Constituent[a]	Mississippi Commercial Hydrated Lime	Airco Carbide Lime
$Ca(OH)_2$	94.3	92.5
$CaCO_3$	2.1	1.85
SiO_3	0.8	1.5
R_2O_3	0.4[b]	1.6
MgO	0.6	0.07
S	0.03	0.15
P	0.01	0.01
Not Analyzed	1.3	2.07

a. Weight percent, based on dry solids.
b. R_2O_3 includes aluminum + iron oxides; analyses by Arthur D. Little indicate 500-1000 ppm Al, which is equivalent to 0.1 to 0.2% Al_2O_3.

Source: LG&E Monthly Progress Report for the period April 1, 1981 to April 30, 1981.

the level of solids in the thickener was decreased. However, not all of the residual carbide lime slurry could be removed before the commercial lime test began. Accordingly, the thickener pH was maintained at less than 10.5 during the first week of the commercial lime test to allow any residual carbide lime to be flushed out of the system before further increases in pH. Aluminum content of the lime feed from the day tank was monitored to confirm that all carbide lime had been removed from the system before the reactor pH was increased. Analyses (dry basis) of lime from the day tank by ADL for 4/7/81 to 4/10/81 gave values of 550-600 ppm Al, indicating that no measurable quantity of carbide lime remained in the system.

Secondary reactor and thickener overflow pH were, nevertheless, increased stepwise rather than immediately to the 11-12 pH design condition because of the remaining potential for aluminum dissolution from fly ash carried over from the ESP. Laboratory tests by ADL with LG&E Unit 6 fly ash indicated that the aluminum in the fly ash is extracted at high pH but not as readily as the aluminum in the carbide lime. The dual alkali system was designed for a fly ash concentration in the inlet flue gas of <0.1 lb/10^6 Btu. The actual inlet concentration of fly ash in the flue gas was measured at 0.85 lb/10^6 Btu during the acceptance test, with approximately 90% removal in the absorber. The fly ash contains about 14% Al_2O_3 — equivalent to approximately 2.0 lb/min Al_2O_3 per scrubber module at 225 MW. By comparison, the aluminum contained in the carbide lime feed at this boiler load is equivalent to 3.04 lb/min Al_2O_3. Thus some potential for aluminum dissolution from fly ash and scaling of the absorber still existed.

Figure 6 shows dissolved aluminum concentration in the thickener overflow as a function of pH from April 7, when the stepped increase in pH was begun, to April 21 when Unit 6 came down due to a boiler tube leak. Secondary reactor pH was rapidly increased from April 7 to April 9, but on April 9 secondary reactor pH was reduced from 10.5-11.5 to 10.0-11.0 because thickener overflow samples sent to Louisville Testing Labs (LTL) indicated that dissolved aluminum might exceed 10 ppm. At the same time, samples sent to METRO Service Laboratories (Louisville) were analyzed as containing about one third the dissolved aluminum analyzed by LTL. Both LTL and METRO were using Atomic Absorption for aluminum analysis. Subsequent analysis by ADL of filtered samples with plasma emission spectroscopy gave better agreement with the METRO aluminum analysis than those done by LTL. (If the samples are not properly filtered, higher values of dissolved aluminum are obtained.)

Because of the uncertainty in aluminum analysis, the pH was not increased again until the last week of testing to minimize potential for aluminum dissolution and scaling.

On April 21 when a boiler tube leak forced shutdown of Unit 6, the two absorbers were opened for inspection. Except for some dried sodium salts, the trays in the "B" absorber module were clean down to the metal. The trays in the "A" absorber were almost as clean, with a light deposit of solids on the top of the bottom tray and a thin transparent deposit on the bottom of the top tray. In general, the trays were extremely clean, and operation for a few days near pH 12 in the thickener (with up to 10 ppm dissolved aluminum) had not resulted in absorber scaling.

7.2 COMMERCIAL LIME TEST RESULTS

7.2.1 Effect of High pH

Because aluminum scaling was a concern early in the commercial lime test and mechanical problems occurred later, the dual alkali system was not operated for a sufficient period at high pH and full load to evaluate the effects of high pH on liquor flow rates and the subsequent longer settling time in the thickener.

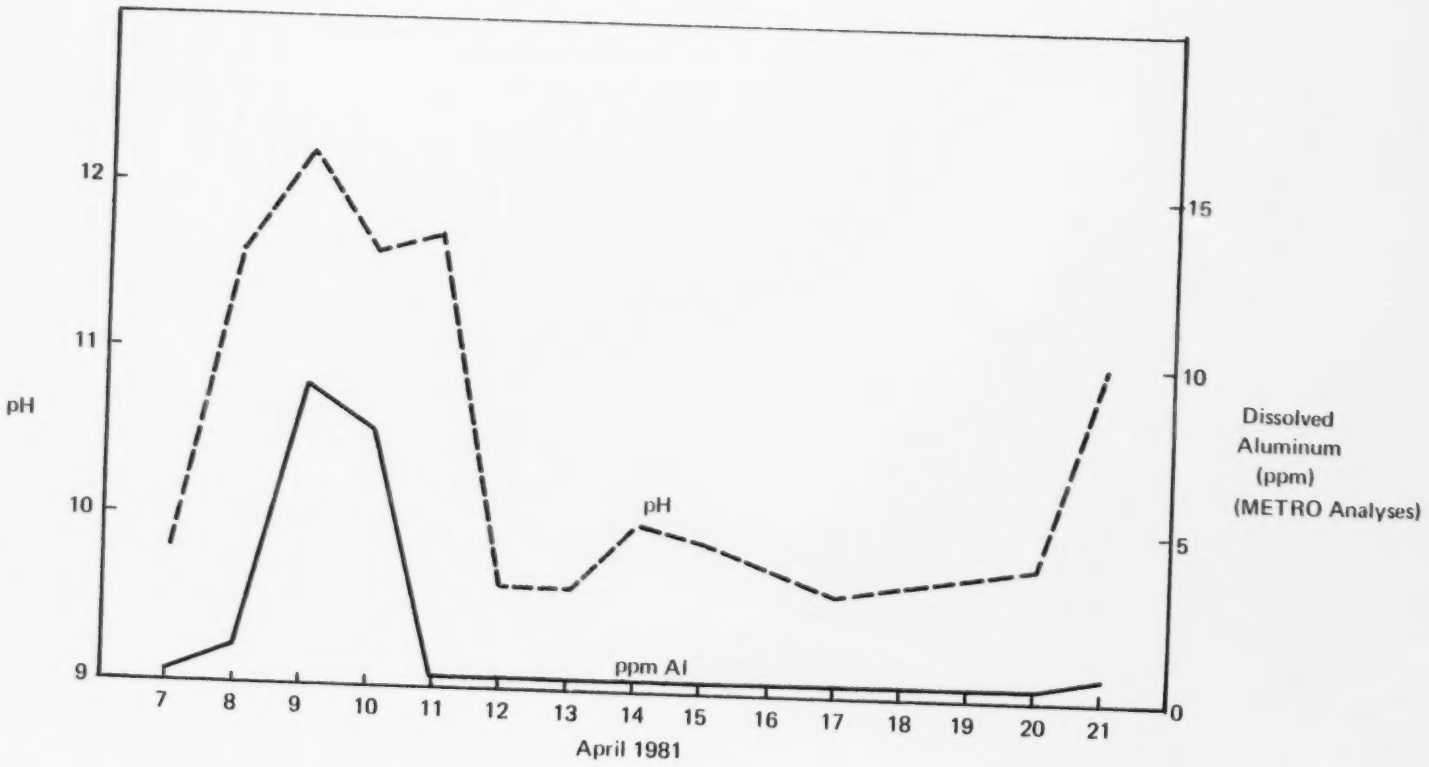

FIGURE 6 RELATION OF pH TO DISSOLVED ALUMINUM IN THICKENER OVERFLOW

7.2.2 Rate and Control of Regeneration Reaction

As expected, the reaction between the commercial lime and scrubber bleed was very rapid. Most of the pH change occurred in the primary reactor, and much less reaction (about 0.5 pH) took place in the thickener. This faster reaction rate together with the higher quality and concentration of lime solids (35 to 40% vs. 20 to 30% for carbide lime) allowed better control of both the reactor system and the water balance.

7.2.3 Reactor Solids Settling

With the exception of one extremely high settling rate measured by Bechtel at the very beginning of the commercial lime testing (>33 ft/hr), the reactor solids settling rates obtained with commercial lime hydrate were not significantly different from those obtained with carbide lime. Table 21 compares typical settling rates of carbide lime reactor solids with those obtained with commercial hydrated lime.

Although most of the settling rates for the solids from the two types of lime are comparable, it is noteworthy that the settling rate in the second settling test with carbide lime (11/5/79), which is at a very high pH, is considerably lower than the average. This low settling rate could result from the very high solids concentration present at that high pH, however. It is also interesting to note that the more reactive commercial lime still produced reasonable settling rates in the higher pH range of 11.0 to 12.0, but this can be accomplished at a much lower solids concentration than with carbide lime.

7.2.4 Filtration of Reactor/Thickener Solids

One potential benefit expected with commercial lime was a higher filter cake solids content, with lower residual sodium because of enhanced cake washing. Unfortunately, there was no significant improvement in the filter cake insoluble solids, even under controlled filter conditions. In addition, the proper cake thickness (1/4 to 3/8 in.) could not be adequately maintained because the higher density solids generated by commercial lime made it difficult for the operators to control the filter tub liquid level, drum speed, and suspended solids concentration to the filter.

Since the proper cake thickness could not be maintained, sodium losses during the commercial lime test did not differ significantly from those encountered with carbide lime. Under controlled filter operating conditions, however, considerably lower levels of residual sodium were possible. This is apparent in Table 22, which compares an earlier filter washing test with carbide lime under controlled operating conditions to the commercial lime filter wash test. As Table 22 shows, both decreased washwater rate and uncontrolled filter cake thickness greatly reduce the ability of the wash to control residual sodium.

TABLE 21

SETTLING RATES OF COMMERCIAL AND CARBIDE LIME REACTORS SOLIDS

Carbide Lime			Commercial Hydrated Lime		
Date	pH	Settling Rate (ft/hr)	Date	pH	Settling Rate (ft/hr)
11/5/79	11.4	13.6	4/7/81	7.4	20.4
	12.1	7.5		11.0	20.4
12/6/79	—	19.7	4/8/81	12.0	17.0
	—	20.4	4/9/81	11.2	17.0
4/16/80	9.5	13.6		11.45	20.4
4/17/80	9.0	17.0	4/10/81	10.95	14.3
				10.95	15.6
7/21/80	7.8	30.6			
	8.1	25.8	4/15/81	9.75	20.4
				9.15	14.3
1/26/81	8.1	17.0			
	8.5	13.6	4/16/81	7.95	12.2
				9.35	13.6
				9.55	16.3
				8.5	19.7
			4/17/81	9.05	17.0
				9.5	12.2
				8.8	20.4

Source: Arthur D. Little, Inc., Settling Tests.

TABLE 22

FILTER WASH TESTS WITH CARBIDE AND COMMERCIAL LIME SOLIDS

Carbide Lime[a]				Commercial Lime[b]			
Cake Thickness (inch)	Wash Rate (gpm)	Cake Loss on Drying (% H_2O)	Percent Sodium in Dry Cake[c]	Cake Thickness (inch)	Wash Rate[d] (gpm)	Cake Loss on Drying (% H_2O)	Percent Sodium in Dry Cake[c]
3/8	32.4	42.3	1.59	3/8	0	44.8	1.86
3/8	44.4	42.0	1.31	3/8	29	43.3	1.12
3/8	64.5	44.1	0.59	3/8	51	46.0	0.42
				3/8	68	49.3	0.13
				1/2	55	46.3	0.93
				9/16	60	48.4	0.34
				3/4	47	48.3	1.68

Notes:

a. Arthur D. Little, Inc., Bechtel, and LG&E analysis performed jointly 6/11/80.
b. Bechtel analysis performed 4/17/81.
c. One and one-half percent residual Na is approximately equal to 0.045 mole Na_2CO_3/mole S in filter cake.
d. At a filter drum speed of 0.60 rpm (3/8-in. cake), 37 gpm of filter cake wash equals one displacement wash.

SECTION 8
STATUS OF LG&E DUAL ALKALI SYSTEM OPERATION

While most of the process, instrument, and mechanical problems of the FGD system have been solved, a few remain. This section discusses the status of those problems that must still be addressed. Background information is available in Section 6.0.

8.1 CONTROL OF pH

The absorber and reactor trains were originally designed to operate on automatic control. With proper cleaning and maintenance of pH electrodes, and the modifications to the reactor pH control system discussed in Section 6.2.4, this should finally be possible.

8.2 LIQUID FLOWMETERS

Although the liquid flowmeters are not critical to proper (automatic) operation of the system, they must be accurate when the initial flow rates to absorber trays and sprays are set. In addition, continued accurate operation is necessary to pinpoint the cause of certain operating problems (such as scaling).

Since the magnetic flowmeters supplied with the system have not proven reliable, they are being replaced with doppler flowmeters, which have proven satisfactory in similar use at Cane Run. These new flowmeters, which will be periodically recalibrated and maintained, should alleviate the problems resulting from inaccurate liquor flow readings.

8.3 PROCESS PIPING

Failures, which continue in the cast fiberglass-reinforced plastic (FRP) piping system, are expected to persist wherever the (FRP) piping has not been replaced. In many areas where the process liquor is not acidic (such as slurry lines and regenerated liquor lines), the FRP piping has already been replaced with carbon steel.

8.4 THICKENER

With repair of the rake torque measuring system and increased awareness of other factors that impair thickener settling efficiency (i.e., sulfate concentration and liquor flow rates), the frequency and severity of thickener upsets should be reduced.

8.5 FILTERS

The filter system, unfortunately, is not capable of automatic operation. Manual control of cake thickness requires *simultaneous* manipulation of filter trough slurry level, drum speed, and thickener underflow solids. This has proven to be a difficult task for the operators, as has maintaining proper wash rate.

Operating the filters on overflow to maintain a constant filter trough slurry depth would, perhaps, have helped the operators to maintain cake thickness. However, the filter overflow line became thoroughly plugged during early operation of the system and was removed.

The difficulties with filter operation have led to a sodium consumption (loss) about twice that of the guarantee. Rosettes of dried salts were observed on the surface of test impoundments of FGD sludge produced even when the filter wash rate and cake thickness were controlled to meet sodium loss guarantee (1% Na by weight in the cake) and treated in the CSI system. The CSI-treated dual alkali sludge impoundments may require capping to prevent sodium-contaminated water runoff. In addition, a more efficient method of washing the FGD wastes may well be required to reduce sodium losses and produce an environmentally acceptable product for disposal.

SECTION 9
REVIEW OF CAPITAL AND OPERATING COSTS

Capital and operating cost estimates for the LG&E dual alkali FGD system were developed during Phase I of the demonstration program. These estimates are summarized in Section 9.1 and are compared with the actual costs incurred at completion of construction and throughout the demonstration program (Sections 9.2 and 9.3).

9.1 ORIGINAL ESTIMATES

The original capital investment and operating cost estimates, summarized below, are described more fully in the Phase I report on preliminary design and cost estimate (EPA-600/7-78-010). The estimated figures for the Cane Run Unit 6, in 1976 dollars, were:

- Capital Investment — $53.2/kW (based on gross peak load of 300 MW) and $57.5/kW (based on gross net load of 277 MW).

- Annual Cost — 2.29 mills/kWh (without reheat system) to 2.65 mills/kWh (with 50F° reheat).

These estimates were based on the following operating conditions: 5,256 operating hours per year, coal sulfur content of 5.0%, SO_2 removal efficiency of 94.2%, and on-site sludge disposal following treatment. (The estimates are subsequently updated in the Phase I report to reflect more accurately the time basis for the capital and operating costs.)

9.1.1 Design Basis

The dual alkali installation at LG&E's Cane Run Station is a retrofit system installed on Unit 6, an existing boiler equipped with an ESP for particulate removal. (The design basis for the system is given in Section 3.) The design gas flow rate of 1,065,000 acfm at the scrubber inlet is about 10% higher than that estimated in an earlier EPA study (EPA-600/2-75-006) for an existing boiler with a capacity of 300 MW. For coal containing 5.0% sulfur and 0.04% chloride (dry basis), the design SO_2 removal efficiency is 94.2%.

The SO_2 removal system has two booster fans, two absorbers, two reactor modules, one thickener, and three filters. The system is designed to operate in the range of 60 to 300 MW boiler load. The waste exhaust gas from the absorber is reheated by hot air from a direct oil-fired reheater. The reheat is equivalent to 50F°. The system has bypass and appropriate spare capacity. The sludge is disposed of in an existing on-site landfill after treatment. The cost of the landfill is not included in the capital cost estimate.

Carbide lime, available from a local supplier, is used as a raw material to regenerate the sodium solution in the dual alkali process. The use of carbide lime rather than commercial lime reduces the operating cost. The carbide lime is supplied in a slurry form by barge and an existing barge handling facility (including a slurry storage tank) is used for the dual alkali system.

9.1.2 Capital Investment

The capital investment originally estimated for the dual alkali demonstration system is shown in Table 23. The overall project schedule was two years, with design starting in the latter part of 1976 and construction to be completed in the latter part of 1978. (Construction was actually completed in early 1979.) The estimated capital investment for the system was $17,379,000 (in 1977 dollars), which corresponded to about $58.0/kW (based on gross peak capacity).

9.1.3 Annual Operating Costs

Since 1979 was planned as the first full year of operation, estimated annual operating costs were developed in 1979 dollars and are shown in Table 24.

Operating costs were based on an average sulfur content in the coal of 3.8%. The average SO_2 removal rate was assumed to be 297 lbs/minute (94.2% SO_2 removal efficiency) at 300 MW boiler load. The operating costs were estimated both with and without operation of the reheat system. When included, reheat was assumed to be 50F°. The sludge disposal costs were based on on-site solids disposal after mixing with fly ash and lime.

The total annual operating cost without reheat was estimated at $4.3 million. The corresponding costs with reheat were about 15% higher. The operating costs range from 2.7 mills/kWh (27.6 cents per 10^6 Btu) without reheat to 3.2 mills/kWh (32.5 cents per 10^6 Btu) with 50F° reheat. These costs included capital charges and depreciation on capital investment.

The cost for carbide lime in 1979 was estimated at $13.29/ton. This included transportation, handling, and preparation of the lime slurry.

9.2 ACTUAL COSTS INCURRED

9.2.1 Capital Costs

Installation of the dual alkali system at Cane Run Unit 6 required capital investment in three different subsystems: FGD, lime slurry feed, and waste processing and disposal. Each subsystem involved independent design and installation efforts, and the associated costs are presented by system in Table 25. More detailed cost breakdowns for the FGD and CSI subsystems are presented in Tables 26 and 27. The costs shown were reported on an as-incurred basis and thus do not represent a constant dollar value of the capital investment.

9.2.2 Operating Costs

Table 28 presents the actual operating cost for the first 12 months of the demonstration period. Table 29 summarizes monthly operating and maintenance costs (less fixed costs of interest and depreciation), and Table 30 shows a sample breakdown of monthly operating and maintenance costs (not including fixed costs of interest and depreciation) for September 1980.

The actual total annual operating cost for the dual alkali system (Table 28) was considerably higher than originally estimated in the Phase I report (EPA-600/7-78-10): $7.8 million vs. $5.1 million. The difference results from greater material and labor requirements than projected as well as substantial increases in the unit cost of raw materials, as discussed below.

9.3 RESOLUTION OF DIFFERENCES BETWEEN ESTIMATED AND ACTUAL COSTS

9.3.1 Capital Investment

The actual capital investment of $19.5 million is somewhat higher than the originally estimated $17.4 million (excluding spare parts and working capital) presented in the January 1978 Phase I report. The material costs for the FGD system and the field supervision were slightly above the original estimate, while total erection labor/overhead and materials for the sludge disposal system were nearly double the original estimates (which were based on a different design that was to involve ponding rather than landfill). Nevertheless, as Table 31 shows, the capital cost for the Cane Run dual alkali FGD system is very close to the reported adjusted capital cost for Southern Indiana Gas and Electric Company's Brown Unit 1 system of comparable size, and less than half the cost of the dual alkali system at Central Illinois Public Service's Newton Unit 1. The capital investments per kW for the LG&E and SIGECO systems are slightly lower than the average investments reported by PEDCo Environmental, Inc. for lime and limestone scrubbing or estimated by TVA for limestone scrubbing (EPA-600/7-82-058a and EPA-600/7-79-177).

9.3.2 Operating Costs

Actual annual operating costs for the dual alkali system were much higher than originally estimated, both in terms of total annual cost and mills per kWh. The estimated total annual operating costs was $5.1 million (with reheat), compared with the $7.8 million actually incurred. The major reasons for this difference are higher labor, maintenance, and soda ash requirements as well as soda ash and fuel oil unit costs that are about twice those originally estimated. Operating cost in mills per kWh is much higher, not only as a result of the increased annual operating cost, but also because the total power generated over the year was equivalent to a load factor of only 44% compared with a projected load factor of 60%. (Table 28 also showed how operation at a higher load factor would affect operating costs.) The FGD system was down for most of the month of December 1980 because of the thickener upset. The combination of operating and maintenance labor costs over this month remained at an average level for the year (since operating personnel were used for some maintenance tasks), but energy output (39,500 MWh) for the month was very low. This month of downtime contributed about 0.2 mills/kWh to the operating costs for the year.

Of the $2.7 million difference between estimated and actual operating costs, $600,000 is for the soda ash makeup requirement — about triple the consumption estimated at about double the unit price. This was partially offset by the fact that less carbide lime was consumed, representing a savings of $100,000 over the demonstration period. Around $400,000 of the differential is due to reheater fuel oil at a unit cost more than double that in the estimate ($1.02 per gallon vs. $0.43 per gallon). Maintenance materials ran 4.1% of total system materials cost compared to the 2.5% used in the Phase I estimate. This accounted for another $240,000. Operating and maintenance requirements were about triple those estimated (155,000 hours vs. 52,000 hours). This represents a $1.3-million difference in operating cost. The labor costs shown in Table 30 are a rough approximation of actual labor cost distribution for the dual alkali system. The greatest requirements for operating labor are about two to three men per shift for the FGD sludge treatment system, about two men per shift for the lime slurry system, with about one to two men per shift each for instrumentation (control room), filter operation and miscellaneous. Total operating labor amounted to an average of about 10 men per shift. Total maintenance labor requirements were equivalent to about eight men per shift, with the largest portion of this requirement for instrument maintenance.

Finally, slightly increased indirect costs for interest and depreciation ($300,000) on the higher capital investment represent most of the remaining difference between estimated and actual operating costs.

Table 32 compares operating costs for the LG&E dual alkali system with those for the other two utility dual alkali systems and those reported by PEDCo for generic lime and limestone systems (EPA-600/ 7-82-058a). The operating cost of the Cane Run dual alkali system at a 60% load factor would still be slightly higher than that reported for the Southern Indiana Gas and Electric Co. (SIGECO) unit (5.8 mills/kWh vs. 5.0 mills/kWh), but considerably less than for the Central Illinois Public Service (CIPS) system.

It should be possible to reduce operating costs further for the LG&E system in two areas, sodium consumption and maintenance. The other two utility dual alkali FGD systems, SIGECO and CIPS, include (or are planning to include) a horizontal belt filter for filtration and multistage washing of the FGD sludge for increased sodium recovery. CIPS incorporated a horizontal belt filter in the original design of its dual alkali FGD system, and SIGECO is replacing its rotary drum vacuum filters with a horizontal belt filter. SIGECO expects to reduce sodium losses by 50%. At LG&E, reduction of sodium losses by 50% would lower operating costs by 0.3 mills/kWh.

Maintenance costs for LG&E's dual alkali FGD system can also be reduced. Table 33 shows a breakdown of maintenance costs experienced during the demonstration period, and a further breakdown of the high maintenance costs items within the various process areas is presented in Table 34.

High maintenance cost items associated with the scrubber modules include the trays, pumps and module tanks. Of these, the recycle pumps have been the highest maintenance cost item as a result of pump lining and impeller failures. Changes made in impeller and pump liner design over the course of the demonstration period should help reduce maintenance costs for these items in the future.

Maintenance costs for instrumentation are nearly equally divided among the five items listed in Table 34 under instrumentation. A major reason for the high maintenance cost for the Lear Siegler stack and du Pont inlet SO_2 analyzers was the frequent calibration required during the demonstration period. Subsequent operation of these analyzers should require less maintenance. Likewise, debugging of the control systems (the panel, recorders and Foxboro electronic control system) accounted for much of the instrument maintenance cost during the demonstration period and less maintenance should be required in the future.

Included with maintenance costs for piping and miscellaneous were costs for insulating and heat tracing of lines. Now that most of this work has been completed, these maintenance costs should also decrease. The cost for repiping, primarily replacement or repair of fiberglass piping, should continue unless these lines are replaced by a more durable piping system.

The maintenance costs for the CSI sludge treatment system are relatively evenly distributed among all the items listed in Table 34, although the mixer, conveyors, and instruments have slightly higher maintenance costs than the other items.

The estimated potential savings for all these high cost areas would total about $386,000 or 0.3 mills/kWh, as shown below:

Area	Present Maintenance Cost ($1000)	Reduced Maintenance Cost ($1000)	Potential Saving ($1000)
Scrubber Modules	132.5 (9.7%)	80.0	52.5
Instrumentation	419.3 (30.7%)	210.0	209.3
Piping & Miscellaneous	310.0 (22.7%)	210.0	100.0
CSI Sludge Treatment	184.4 (13.5%)	180.0	4.4
Other	319.7 (23.4%)	300.0	19.7
Total	1,365.9 (100%)	980.0	385.9

Maintenance cost reduction, together with improved sodium recovery and a unit operating load factor of 60%, could lower the operating cost for the LG&E dual alkali system from the 6.9 mills/kWh actually experienced to about 5.0 mills/kWh, much closer to that reported for SIGECO (EPA-600/7-82-058a).

TABLE 23

ESTIMATED CAPITAL COSTS FOR DUAL ALKALI FGD SYSTEM AT CANE RUN UNIT 6 ($ 1977)

	Dollars
Total Materials:	
Major Equipment	7,037,000
Other Materials	2,525,000
Sludge Disposal Equipment[a]	900,000
Additive Slurry System[b]	700,000
Total Materials Cost	11,162,000
Erection:[c]	
Direct Labor (252,800 hrs. @ $12/hr)[d]	3,034,000
Field Supervision	273,000
Total Erection Cost	3,307,000
Engineering Costs:	
System Supplier Engineering	1,323,000
Owner's Engineering Expenses	303,000
Owner's Consulting Engineer	852,000
Total Engineering Cost	2,478,000
Spare Parts, 2% of the Total Materials Cost	232,000
Working Capital	200,000
Total Capital Investment:[c]	17,379,000
$/kW (based on 300 MW gross peak load)[e]	57.9
$/kW (based on 277 MW gross net load)	62.7

Basis: 300 MW gross peak load existing coal-fired boiler (277 MW net peak load). Coal sulfur content, 5.0%. Scrubber SO_2 removal efficiency, 94.2%. Stack gas reheat, 50°F. Disposal of sludge (after treatment) to on-site landfill. Project beginning mid 1976, ending late 1978, average 1977 dollars. Necessary parts in storage and reasonable spare capacity.

Notes:

a. Sludge disposal equipment cost ($900,000), shown separately, is the 3/7 portion of the total sludge disposal cost for Cane Run Units 4, 5 and 6. It does not cover the CSI sludge treatment system.
b. Additive supply system cost ($700,000), shown separately, is the 3/7 portion of the total carbide lime additive supply system for SO_2 removal systems for Cane Run Units 4, 5 and 6.
c. Erection equipment cost is included in plant overhead.
d. Includes plant overhead.
e. The capital investment for the above system in 1976 dollars was equivalent to $15.95 million or 53.15 $/kW (based on gross peak capacity).

Source: EPA 600/7-78-010.

TABLE 24

ESTIMATED AVERAGE ANNUAL OPERATING COSTS FOR DUAL ALKALI FGD AT CANE RUN UNIT 6 ($ 1979)

Direct Costs;	**Quantity**	**Unit Cost, $**	**Total Costs, $**
Carbide Lime	58,728 ton	13.29/ton	780,500
Soda Ash	1,912 ton	78.65/ton	150,400
Fuel Oil	1,802,808 gal.	0.43/gal.	775,200
Electricity	6,188,480 kWh	0.01/kWh	161,900
Water	126,100,000 gal.	0.05/1,000 gal.	6,300
Sludge Removal	185,280 ton	2.01/wet ton	372,400
Maintenance Materials	2.5% of total materials cost		279,000
Labor			
Operation	26,280 hr	8.18/hr	215,000
Maintenance	26,280 hr	8.28/hr	217,600
Analysis	2,080 hr	10.00/hr	20,800
Supervision	2,080 hr	various	40,000
Total Direct Costs			3,019,100
Indirect Costs:			
Overhead	59.4% of 493,368 (Total Labor)		293,100
Interest	6.125% of 17,379,000		1,064,500
Depreciation	4.17% of 17,379,000		724,700
Total Indirect Costs			2,082,300

	With Reheat	**No Reheat**
Total Annual Operating Cost, $	5,101,400	4,326,200
Mills/kWh	3.24	2.74
¢/10^6 Btu	32.5	27.6
$/Ton of S Removed	217.9	184.7

Basis: 300 MW (gross peak capacity) existing coal fired plant, 60% load factor, 9,960 Btu/kWh, 3.8% S in coal 94.2% SO_2 removal, average SO_2 removed — 297 lbs/min. at 300 MW (gross peak) load, on-site solids disposal by trucking after treatment, stack gas reheat 50F°.

Source: EPA-600/7-78-010.

TABLE 25

ACTUAL CAPITAL COST BREAKDOWN BY SUBSYSTEM FOR CANE RUN UNIT 6
($ 1977 - 1979)

Subsystem	Material Costs[b]	Erection Costs			Engineering Costs[a]		
		Direct Labor	Field Supervision and Engineering	Construction Overhead	System Supplier	LG&E Consultant	Total
FGD	10,183,200	2,386,400	277,900	1,373,000	833,500	670,000	15,724,000
Lime Slurry[c]	460,600	118,300	9,100	112,000	—	—[d]	700,000
Waste Disposal	2,084,200	387,400	32,500	398,700	208,400	—[d]	3,111,200
Total	12,728,000	2,892,100	319,500	1,883,700	1,041,900	670,000	19,535,200

a. Owner's engineering is included in field supervision and engineering.
b. Includes spare parts at 1% of total costs.
c. Reported as a portion of the overall lime slurry system for Units 4, 5 and 6.
d. Included in field supervision and engineering.

TABLE 26

MATERIALS COSTS FOR THE FGD SYSTEM*

Process Materials	
Process Equipment:	
— Absorbers	909,600
— Tanks	615,100
— Thickener Shell and Mechanism	343,300
— Vacuum Filters	980,000
— Reheaters	160,200
— Fans and Fluid Drives	575,400
— Pumps and Motors	842,800
— Agitators	96,800
— Weigh Feeder	50,400
— Lining	428,600
Total Process Materials	5,002,200
Process Instrumentation	443,000
Process Piping, Insulation, and Heat Tracing	640,300
Other Process Materials:	
— Ductwork and Dampers	1,047,000
— Structural Steel	326,500
Total Instrumentation	2,456,800
Additional Materials	
Utility and Service Piping and Instrumentation	
— Instrumentation	74,000
— Piping, Insulation, and Heat Tracing	87,000
— Miscellaneous	70,000
Total Utility and Service P&I	231,000
Electrical	
— Electrical Auxiliaries (4160, 460, 220/110 volt)	491,200
— Station Grounding and Ducts	38,400
— Station Control (wiring to instruments)	207,600
Total Electrical	737,200
Other	
— Foundations (includes lagging and insulation for fans)	973,400
— Buildings	222,200
— Land Improvements, Roads, and Storm Runoff Containment	560,400
Total Other	1,756,000
Total	10,183,200

*As-incurred plus estimate-for-completion basis.

TABLE 27

ACTUAL CAPITAL COST BREAKDOWN FOR WASTE DISPOSAL SYSTEM
($ 1977-1979)

		Erection Costs		Engineering Costs	
	Material Costs	Direct Labor	Field Supervision & Engineering	Construction Overhead	System Supplier
Waste Disposal System[a]					
Conversion Systems, Inc.	1,673,100	385,900	32,400	398,400	—[b]
Landfill	411,100	1,500	100	300	—[b]
Total	2,084,200	387,400	32,500	398,700	208,400

Total Cost for Waste Disposal System: $3,111,200

a. Does not include the pneumatic fly ash conveyor system, which was installed in 1981 to service Cane Run Units 4, 5 and 6 at a capital cost of approximately $200,000.
b. 10% of material cost.

TABLE 28

ACTUAL AVERAGE ANNUAL OPERATING COSTS (WITH REHEAT) FOR CANE RUN UNIT 6 DUAL ALKALI FGD SYSTEM (May 1, 1980 — April 30, 1981)

Direct Costs:	Quantity	Unit Cost, $	Total Costs, $
Carbide Lime	44,647 ton	15.32/ton	683,992
Soda Ash	5,438 ton	134.53/ton	731,574
Fuel Oil	1,204,279 gal.	1.02/gal.	1,228,365
Electricity	12,925,000 kWh	0.012/kWh	155,100
Water	79,170,000 gal.	0.05/1,000 gal.	3,959
Sludge (Trucking & Landfill)	117,890 ton	2.50/wet ton	294,725
Maintenance Materials			506,280
Labor			
Operation	83,670 hrs	8.45/hr	707,012
Maintenance	71,820 hrs	8.55/hr	614,061
Analysis	2,080 hrs	12.00/hr	24,960
Supervision	2,080 hrs	25.00/hr	52,000
Total Direct Costs			5,002,028
Indirect Costs:			
Overhead (Payroll & Plant)	52.9% of 1,398,033 (Total Labor)		739,560
Interest	6.125% of 19,535,200		1,196,531
Depreciation	4.65% of 19,535,200		908,387
Total Indirect Costs			2,844,478
Total Annual Operating Cost, $			7,846,506
Actual Operating Cost (at 44% load factor)			
Mills/kWh			6.9
¢/10^6 Btu			69.2
$/ton of S Removed			492.5
Adjusted Operating Cost (projected for 60% load factor)			
Mills/kWh			5.8
¢/10^6 Btu			58.1
$/ton of S removed			411.3

Basis: 300 MW (gross peak capacity) existing coal-fired plant, 9,960 Btu/kWh, 3.5% S coal, 91.6% SO_2 removal, average SO_2 removed — 279 lbs/min. at 300 MW (gross peak) load, on-site solids disposal by trucking after treatment, stack gas reheat 50F°.

TABLE 29

LG&E UNIT 6: SUMMARY OF OPERATING AND MAINTENANCE COSTS FOR DUAL ALKALI DEMONSTRATION (Phase IV)

	Unit 6 Gross Power (MWH)	Operating Cost* ($1000)	Maintenance Cost ($1000)	Operating Cost* (mills/kWh)	Maintenance Cost (mills/kWh)
May 1980	118,040	352.7	157.7	3.0	1.3
June	108,040	402.5	160.1	3.7	1.5
July	109,136	469.5	32.2	4.3	0.3
August	108,742	362.0	170.1	3.3	0.9
September	104,772	325.2	103.1	3.1	0.9
October	74,467	327.6	140.5	4.4	1.9
November	55,406	243.2	107.8	4.4	1.9
December	39,531	154.8	130.2	3.9	3.2
January 1981	147,934	273.8	153.2	1.8	1.0
February	75,150	397.4	89.6	4.0	1.2
March	97,707	336.5	69.6	3.4	0.7
April	94,251	422.2	51.8	4.5	0.6
Total	1,133,176	4,067.4	1,365.9	Avg. 3.7	Avg. 1.3

*Does not include maintenance costs, plant overhead, interest or depreciation.

TABLE 30

OPERATING AND MAINTENANCE COSTS FOR CANE RUN UNIT 6 DUAL ALKALI SYSTEM SEPTEMBER 1980[a]

	Operating			Maintenance		
	Labor	Other	Total	Labor	Parts	Total
Lime Slurry System	18,110	57,680[b]	75,790	1,890	1,630	3,520
Soda Ash System	1,360	66,520[c]	67,880	670	2,780	3,450
Scrubber Modules	1,850	—	1,850	9,120	7,460	16,580
Gas Side Equipment	2,020	99,800[d]	101,820	1,100	1,420	2,520
Reactors	70	—	70	—	(40)	(40)
Thickener	3,570	—	3,570	2,300	2,620	4,920
Vacuum Filters	12,450	1,660	14,110	2,910	1,620	4,530
Instrumentation	9,360	—	9,360	17,560	3,470	21,030
CSI	22,500[e]	9,790	32,290	9,360	11,230	20,590
Service Water	—	—	—	—	—	—
Piping & Misc.	10,030	8,440	18,470[f,g]	21,010	4,970	25,980
Total			325,200			103,080
Mills per kWh			3.1			1.0

Notes:

a. Unit 6 Gross Power Generated September 1980: 104,772 MWh
b. Airco & McBride (carbide lime)
c. Stauffer Chem. (soda ash)
d. Reheater Fuel Oil
e. Incl. $11,360 Eng. & Supervision
f. Includes Sludge Hauling of $4,770
g. Includes $6,080 Laboratory Costs

TABLE 31

COMPARISON OF CAPITAL COSTS FOR FGD SYSTEMS

FGD System	Size (MW)	Utility	Capital Cost ($/kW) Reported	Adjusted
Dual Alkali[a]	265	Southern Indiana Gas & Electric Co.	47.2 (1978 $)	88.4 (1981 $)
Dual Alkali[a]	300	Louisville Gas & Electric Co.	71.5 (1978 $)	87.8 (1981 $)
Dual Alkali[a]	617	Central Illinois Public Service	174.8 (1978 $)	263.9 (1981 $)
Lime[b]	—	—	81.8 (1978 $)	116.5 (1981 $)
Limestone[b]	—	—	67.9 (1978 $)	98.9 (1981 $)
Dual Alkali[c]	500	—	103.59 (1979 $)	—
Limestone[c]	500	—	79.75 (1979 $)	151 (1982 $)[d]

Notes:

a. Reported by PEDCo Environmental, Inc. (EPA-600/7-82-058a). Reported costs obtained from Edison Electric Institute and by direct contact with utilities. Costs associated with particulate control were then deducted; modifications necessitated by the FGD system were added, and the resulting capital cost was escalated to 1981 dollars and presented as the "adjusted cost."

b. Reported by PEDCo Environmental, Inc. (EPA-600/7-82-058a). Cost figures shown represent average costs for installations ranging in size from 60 to 920 MW (lime FGD systems) and 100 to 900 MW (limestone systems).

c. Estimated by TVA for generic system (EPA-600/7-79-177, Tables A-7 and 8-40; Fixed Investment Divided by Gross MW).

d. Estimated by TVA for generic system ("FGD Economics in 1980" in EPA-600/9-81-019a, p. 78; Fixed Investment Divided by Gross MW).

TABLE 32

COMPARISON OF OPERATING COSTS FOR FGD SYSTEMS

FGD System	Size (MW)	Utility	Capital Cost ($/kW) Reported	Adjusted
Dual Alkali[a]	265	Southern Indiana Gas & Electric Co.	1.3 (1978 $)	5.0 (1981 $)
Dual Alkali[a]	300	Louisville Gas & Electric Co.	—	7.2 (1981 $)
Dual Alkali[a]	617	Central Illinois Public Service	—	13.9 (1981 $)
Lime[b]	—	—	3.2 (1978 $)	8.1 (1981 $)
Limestone[b]	—	—	1.6 (1978 $)	6.1 (1981 $)
Dual Alkali[c]	500	—	3.1 (1980 $)	—
Limestone[c]	500	—	3.0 (1980 $)	8.42 (1984 $)[d]

Notes

a. Reported by PEDCo Environmental, Inc. (EPA-600/7-82-058a). Reported costs obtained from Edison Electric Institute and by direct contact with utilities. All non-labor costs were adjusted to a common 65% capacity factor. Direct costs, overhead, and fixed costs that were not reported were estimated and added. All annual costs were escalated to 1981 dollars and presented as the "adjusted" costs.

b. Reported by PEDCo Environmental, Inc. (EPA-600/7-82-058a). Cost figures shown represent average costs for installations ranging in size from 60 to 920 MW (lime FGD systems) and 100 to 900 MW (limestone systems).

c. Estimated by TVA for generic system (EPA-600/7-79-177, Tables A-8 and A-41; mills/kWh less cost of capital and taxes).

d. Estimated by TVA for generic system ("FGD Economics in 1980" in EPA-600/9-81-019a, p. 79; first year revenue requirements less capital charges for land, working capital, and interest during construction and start-up).

TABLE 33

CANE RUN UNIT 6 DUAL ALKALI FGD SYSTEM: MAINTENANCE COST BREAKDOWN FOR DEMONSTRATION PERIOD

	% Annual Maintenance Cost
Carbide Lime Supply System	4.6
Soda Ash System	1.4
Scrubber Modules	9.7
Gas Side Equipment	4.7
Reactors	1.0
Thickener	4.9
Vacuum Filters	6.0
Instrumentation	30.7
CSI Sludge Treatment System	13.5
Service Water	0.8
Piping & Miscellaneous	22.7
Total	100.0

TABLE 34

ITEMIZATION OF HIGH MAINTENANCE COST ITEMS

Scrubber Modules (9.7% of annual maintenance cost)
- Trays
- Pumps
- Module Tanks

Instrumentation (30.7% of annual maintenance cost)
- Stack Continuous Emission Monitor for O_2 and SO_2
- Inlet SO_2 Analyzer
- FGD System Control Panel
- Recorders, Gauges
- Electronic Control System

Piping & Miscellaneous (22.7% of annual maintenance cost)
- Insulation
- Heat Trace
- Repiping

CSI Sludge Treatment System (13.5% of annual maintenance cost)
- Lime Handling
- Fly Ash Handling
- Filter Cake
- Pug Mill Mixer
- Mixer Conveyor
- Stacker Conveyor
- Dust Collector
- Instruments
- Sump

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